ESTIMATION OF SPECIFIC DEPOSIT IN GRANULAR WATER FILTER

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CERTIFICATE

Certified that the work presented in this thesis entitled "Estimation of Specific Deposit in Granular Water Filter" by Sri Prasanta Kumar Bhunia has been carried out under my supervision and it has not been submitted elsewhere for a degree.

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From studies reported in the literature it is apparen that the physicochemical phenomena involved in filtration of water through granular media are rather complex and more understanding is required for efficient design and effective operational control of water filtration plants. The objective of present study was to (i) investigate the nature of specific deposit change with filtration time and filter depth using several media-suspension combination, e.g., sand and coal, and kaolinite turbidity, alum coagulated kaolinite turbidity and wastewater, and (ii) develop a possible relationship between the _in situ_ specific deposit estimated through Fair-Hatch modification of Kozney-Carman equation and that obtained using backwash characteristics or influent (primary) suspension characteristics along with filter performance data.

It is observed that for a particular media-suspension combination, specific deposit is proportional to flow rate and the net solids uptake (difference between average influent and effluent turbidity).

Specific deposit build up is faster in the case of sand compared to coal and its rate increases in the order kaolinite, kaolinite (coagulated), and wastewater.

Based on observations reported, specific deposit estimation by backwash approach appears to be a feasible technique. In case of a sand or coal-sand dual-media filter receiving alum coagulated influent, specific deposit (backwash) should be divided by a factor 2.5 to obtain *in situ* specific deposit. But in case of wastewater filtration through sand or coal-sand dual-media filter, specific deposit (backwash) probably parallels *in situ* specific deposit. For predicting headloss during filtration, Shektman's (1961) model may be employed.

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LIST OF SYMBOLS

A	=	Hamaker's constant
C	=	Local concentration of the particles in the flowing liquid, mg/l
C_o	=	Effluent concentration, mg/l
c	=	Constant
D	=	Diffusion coefficient
D_1	=	Dielectric constant
d_1, d_2	=	Diameter of solid particles, cm
d_e	=	Diameter of equivalent channel, cm
d_m	=	Diameter of media grain, cm
d_p	=	Diameter of suspended particle, cm
e	=	Stokes particle diameter, cm
E	=	New filter coefficient
E'	=	Streaming potential, mV
F	=	Attractive force
f	=	Porosity
f_o	=	Clean bed porosity
f_t	=	Porosity at any time t during filtration
G	=	Emperical Constant in Deb's equation
G_1	=	Velocity gradient
g	=	Acceleration due to gravity, cm/sec^2

H = Headloss, cm

H = Pressure drop across the filter

H_o = Headloss in a clean filter bed, cm

I = Interception parameter

Js_o^2 = Filter media geometry characteristics

K' = Boltzmann's constant

K_o = Carman shape factor

K_e = Electrical conductivity

K'' = Coefficient of permeability

K_1, K_2, K_3, K^1 and K''' = Constants

L = Distance from the inlet face of the filter media, cm

L_e = Length of the actual flow path of the media, cm

m = Hydraulic radius, cm

N = Intertial parameter

n = Number, atoms/cm^3

N_o = No. of suspended matter originally present, nos./ml

N_1 = Concentration of particles, nos. /ml

N_t = Remaining free unflocculated matter, nos./ml

P_e = Peclet number

p = Constant depending upon the grain size

r = Distance between the two spheres, cm

r_1 = Area volume ratio of coated sphere

S	=	Specific surface area
T	=	Temperature in degrees absolute
U_o	=	Characteristic flow velocity
V	=	Rate of filtration, lpm
V_o	=	Superficial velocity
V_t	=	Volume of media in the filter column, cm^3
v	=	Floc volume concentration
W_s	=	Weight of the media, gm
x,y,z	=	Constants
η	=	Transport efficiency of an individual grain
n_D	=	Combined convective diffusion
ϕ	=	Sphericity
γ	=	Sticking ratio
λ	=	Filter coefficient, cm^{-1}
λ'	=	Modified filter coefficient
λ_o	=	Clean bed filter coefficient
μ	=	Viscosity, poise
ρ_1	=	Density of fluid, gm/cm^3
ρ_p	=	Density of the particle, gm/cm^3
γ_s	=	Specific weight
ν	=	Kinematic viscosity, poise
σ	=	Specific deposit
σ_u	=	Ultimate or saturation value of specific deposit
ζ_m	=	Zeta potential of filter media, mV
ζ_p	=	Zeta potential of suspended particle, mV
ξ^2	=	Constant

1. INTRODUCTION

Filtration is a natural phenomenon which has been in use in municipal water treatment under engineered condition. The basic objective of water filtration is to produce a low turbidity water which is aesthetically acceptable and suitable for disinfection. More recently, there has been an increasing use of granular filtration in physico-chemical treatment of wastewater. Present understanding of the filtration phenomena does not allow the prediction of filter performance based solely on theoretical considerations. Consequently, various mathematical models based on theoretical considerations as well as experimental data have been proposed to predict filter performance in terms of head loss development which happens to be a major operational control parameter. Inspite of the limitations inherent to these models they are practically useful both in terms of design and operational control. All these equations make use of the term "specific deposit" (volume of particulate matter deposited per unit volume of the filter bed) which needs to be estimated to predict the headloss development. Several investigators have also indicated the significance of specific deposit in predicting the filter efficiency (filter coefficient).

Estimation of specific deposit or floc volume

concentration in a filter has been attempted by several investigators using various approaches. These approaches include centrifuging the flocculated influent suspension, measurement of the distribution of radioactively levelled algae with a scintillation counter, timing ultrasonic pulse across a clean filter bed and one clogged with quartz powder, and backwashing of the filter and collection of the backwash waste in an Imhoff Cone. However a systematic study to investigate the variation in specific deposit with depth and time of the filtration as well as with filter media-suspension combination would lead to more meaningful application of this parameter.

The present research was undertaken to study in details the specific deposit change in a filter for several media-suspension combinations. Specific deposit values were computed using the settled sludge volume concentration of primary (influent) suspension and backwash waste, and the change in filter bed porosity. One of the objective was to standardize a simple, acceptable method for its estimation from primary or backwash waste characteristics.

2. PRESENT WATER FILTRATION TECHNOLOGY AND THEORETICAL BACKGROUND

2.1 Filtration Practice

The basic objective of water filtration is to remove from water the suspended and colloidal particulate matter constituting turbidity so as to produce an aesthetically acceptable low turbidity effluent which is suitable for disinfection. The granular filters commonly employed in water treatment are slow and rapid sand filters and more recently the dual-media filter or multi-media filters. Pressure filters employing sand or diatomaceous earth are often used in small installations.

2.1.1 Slow Sand Filter

This is the earliest and simplest type of water filter with no requirement of pretreatment or mechanical appliances and more or less resembles filtration through natural ground strata. The filtration rate is very low and therefore is suitable in rural areas where plenty of land is available at cheap rate. Huisman (1970) has described in details the design procedures and suggested methods for their effective operation and control. Fine sand having effective size of 0.25 to 0.35 mm and uniformity coefficient of 2 to 3

is the granular media used which enhances simple mechanical straining of the suspended particles. Due to straining the impurities accumulate on the surface of the filter in the form of a mat called 'schmutzedecke' which is also biologically active and is considered to be the most significant feature in the removal of suspended impurities. For effective filtration, these filters are operated at an exceedingly low rate of about 2 lpm/m^2. Slow sand filters are highly efficient in the removal of bacteria in raw water (98% - 99%) when it is not overloaded but it may be insufficient to prevent the passage of pathogenic bacteria so as to assure complete safety against water borne disease. These filters have some beneficial effect on odours and tastes particularly due to algae or suspended matters. Agarwal (1973) brought out a falacy in preventing pretreatment to water applied to these filters and showed that the filtration rate can be increased upto 18 lpm/m^2 using pretreatment (alum coagulation and settling).

2.1.2 Rapid Sand Filter

In rapid sand filters, coarser and more uniform sand media having effective size of 0.45 to 0.6 mm and uniformity coefficient of 1.5 and lower is used and these filters are provided with mechanical control equipments.

The distinguishingfeatures of the rapid sand filters are their relatively small size and higher filtration rate, mandatory pretreatment (chemical coagulation and settling), and backwash requirement as impurities accumulate at nearly all depths. Fair et al. (1968) discussed the design of these filters in details. The rapid sand filter is designed for a filtration rate of 73.5 to 98 lpm/m^2 and the filter run is terminated when the headloss through the filter reaches about 250 cm. The filter is then backwashed by upflow of wash water with or without auxilliary scouring and the amount of washwater used is 4% to 6% of the water filtered. Stratification of the bed during backwashing is the major drawback in the rapid sand filters. In the conventional single media rapid sand filter, the sand particles are hydraulically graded during backwashing with the result that the finest particle is at the top and the coarsest at the bottom. As the pore opening of the sand in the top of rapid sand filter is less, the most of the suspended particle gets arrested in top few centimeters of the filter bed and the filter gets clogged much earlier with consequent increase in the headloss. Hence, the filter is required to be backwashed although the lower layers are still capable of removing suspended impurities.

2.1.3 Multimedia Filter

A recent advance in the filtration technology which allows much higher filtration rates, more effective utilization of the filter bed is the multi-media or multi layer filtration. In these filters the granular media is both size and density graded so that the coarsest, least-dense medium is at the top and the finest, densest medium is at the bottom (Mohanka, 1969).Use of large particles above the smaller, more dense particles results in "in depth" filtration which promotes more efficient use of the bed depth allowing higher loading rates. The multi-media filter commonly employs a tri-media bed consisting of anthracite (Sp. gravity 1.5), sand (Sp. gravity : 2.6) and garnet (Sp. gravity 4.5) as well as coal-sand dual-media and is generally designed for a filtration rate of 205 to 330 lpm/m^2. The United States Environmental Protection Agency (EPA) design guidelines (Oct. 1971), state that it is desirable to have as coarse a media as possible in the top to prevent surface binding and then sand and the garnet as fine as possible to promote high degrees of removals. However, the disparity in sizes cannot be too great or intermixing of the coal by sand or sand by garnet will result after backwashing.

2.2 Theory of Filtration

Iwasaki (1937) based on his observation, proposed that the change in concentration of suspended particles per unit depth in a column of granular filtering media is proportional to the local concentration of the particles in suspension;

$$- \frac{\partial C}{\partial L} = \lambda C \qquad (1)$$

where, C = local concentration of particles in the flowing liquid,

L = distance from the inlet face of the filter media, and

λ = filter coefficient which is a measure of the efficiency of the filter.

Based on the theory of the filtration of aerosole particles through fibrous filters, Agarwal (1966) showed theoretical justification for this equation. The assumptions in such derivation is that the total removal is contributed by individual fibres (or in case of granular filters by individual grains) and an individual grain totally removes all particles from a certain percentage of its cross-sectional area confronting the flow. The filter coefficient, λ, for such air filter can be computed by determining the number of grains in the unit depth of the filter media and

the removal efficiency of each grain as follows :

$$\lambda = \frac{1.5\ (1-f_o)\,\eta}{d_m} \tag{2}$$

where, η = efficiency of an individual grain,

f_o = porosity of the clean filter, and

d_m = diameter of the filter grain.

Iwasaki (1937) also stated that the volume of solids removed from water is volume deposited in the bed :

$$\frac{\partial \sigma}{\partial t} = - V \frac{\partial C}{\partial l} \tag{3}$$

where, $\frac{\partial \sigma}{\partial t}$ = time rate of change of the deposit ratio at a particular depth l and time t,

$\frac{\partial C}{\partial l}$ = rate of decrease of volumetric concentration of the floc in water,

V = rate of filtration, lpm, and

σ = specific deposit or deposit ratio.

According to Camp (1964), equation (3) is valid regardless of the manner of deposit within the bed, but its validity is restricted to the cases in which the volume of floc already deposited is not reduced by loss of water. Subject to this restriction, if σ is known as a function of time during a run, equation (3) can be used to compute $\frac{\partial C}{\partial l}$ at a particular

depth and time.

Deb (1969) contends that Iwasaki's relationship is over simplified, in that it does not consider local suspension concentration gradients and deposit concentration with respect to time and depth. He modified Iwasaki's relationship to the form :

$$-\frac{\partial C}{\partial l} = \frac{1}{V} \frac{\partial \sigma'}{\partial t} + \frac{f_o - \sigma'}{V} \frac{\partial C}{\partial t} \tag{4}$$

where, f_o = porosity of the clean bed.

Deb (1969) introduced time as an additional rate parameter and described the relationship as total differential :

$$\frac{DC}{Dt} = V \frac{dC}{dl} + \frac{dC}{dt} = -\lambda' C \tag{5}$$

where, λ' = modified filter coefficient, sec^{-1}

Patwardhan (1973) suggested a new filter coefficient, E which is defined as the fraction of suspension concentration removed at each contact or at each grain across the flow direction. The consideration of unit steplength as long as grain diameter does not change seem to be appropriate as the removal at one grain would be a random phenomena and may not represent the removal in the uniform granular media. Average value of E avoids such difficulty and can be used advantageously for the multi-media filters, but as the

coefficient, λ_o is used by most of the investigators, its use will be preferable for only comparative studies.

As the filtration is basically a physicochemical process, all relevant parameters affect the value of λ and make its prediction difficult. The size, shape and the nature of the media, suspended floc particles, operational parameters like filtration rate, influent water characteristics like pH, alkalinity, specific ion concentration and viscosity. All influence the filter efficiency and their combined effect is supposed to be included in the value of λ. Moreover as water flows through the media λ goes on changing during the filter run due to deposition of suspended particles and subsequent change in the media property. The initial clean bed filter coefficient, λ_o, indicates the filter efficiency under clean bed condition where media structures and properties can be determined with somewhat more accuracy). The particle removal mechanisms developed for explaining filtration phenomenon are therefore based on relation between various parameters and the λ_c values.

2.3 Removal Mechanisms

O'Melia and Stumm (1967) conceived filtration as a two step process, the first being the transport of particles from the bulk of water in the pore space to the grain and the

second and subsequent step being the attachment of particles to the grain surface or to the particles previously deposited. Particle transport is a physical process and is principally affected by physical characteristics of the filter bed, its method of operation and the suspension under treatment. Particle attachment depends upon the chemical characteristics of the aqueous phase and the surface characteristics of the suspended particles as well as the filter media. So, particle attachment is a chemical process and is influenced by both physical and chemical parameters.

2.3.1 Mechanical Straining

This is the simplest and one of the most popular mechanisms proposed by Hazen (1904), Fair and Geyer (1954), Hall (1957), Camp (1964) and Agrawal (1966). This could be considered in fact as a complete removal mechanism and is probably not affected by the variations in attachment conditions. Accordingly, a filter bed is considered to be a set of sieves placed one below another with opening sizes equal to pore sizes which are of same order of grain sizes. Thus, the pores in the bed are considered to be two dimensional instead of considering them as a continuous capillary channels. If the size of the particles in suspension or its any one dimension is larger, than the pores, it is likely to be

arrested. Removal by this phenomena depends on the size of media grains and the particle size distribution of the suspension and is not affected by any other physicochemical parameters or operating conditions. The particles being removed is in general smaller than the pores, the removals could be mainly at the narrow corners of the interstices. According to Ives (1971), straining can take place when the concentration of particles is too high in which case many particles arrive simultaneously at a pore opening and jam in it by arching action.

An extended form of straining was observed by Sakthivadivel (1972) and he referred to the phenomenon as bridging. The results of his experiments indicate that interstitial straining and bridging are the principle mechanisms of removal of non-colloidal particles. The critical parameter which controls the clogging or non-clogging of a porous media is the ratio of the diameter of the grain to that of a suspended particle and the type of packing for a given porosity. If the ratio of the diameter of the medium to the diameter of the particles in suspension is 5.0, the filter would clog regardless of the porosity. The ratio above which the suspended particles would not be deposited is 14.0. Hall (1957) pointed out that the area of the

narrow corners of crevices which would be very much smaller than the total pore area showed considered effective in the removal by the straining phenomenon. Based on Hall's expression, Agrawal proposed the following equation :

$$(\lambda_o)_{st} = 3.5\ d_p^{3/2}\ \ dm^{-5/2} \tag{6}$$

where, $(\lambda_o)_{st}$ = filter coefficient of the clean filter due to straining, and

d_p = diameter of the particle.

Hall (1957) assumed that the suspended particles are uniformly distributed over the entire cross-section and absence of any viscous forces in the flow regime which are not justifiable. Agrawal (1966) drew attention to laminar distribution and diversion of major flow away from the corner spaces, which suggests that straining alone cannot explain the removal of turbidity in filters.

2.3.2 Transport Mechanism

Ives (1970) proposed that there are several transport mechanisms which probably act simultaneously but their relative importance depends on the characteristics of the suspended particles (principally size, density and shape), the characteristics of flow (principally velocity, viscosity and temperature), and the characteristics of filter

media (principally surface area, pore size, shape and volume).

2.3.2.1 Diffusion

The transport of very small particles in a moving fluid occurs due to convective diffusion which is the combination of two different mechanisms. First is the Brownian random motion of particles due to the thermal energy of surrounding water molecules and the second is due to entrainment of particles in convective flux of flow. Levich (1962) developed an equation for determining single collector efficiency (η_D) with Peclet number (P_e) as dimensionless parameter for combined convective diffusion :

$$\eta_D = 4.04\ P_e^{-2/3} \tag{7}$$

in which, $P_e = \frac{U_o L}{D}$

where, U_o = characteristic flow velocity,

L = characteristic length along which the major change in concentration takes place, and

D = diffusion coefficient

Small Peclet number indicates concentration distribution largely by molecular diffusion. Such situation occurs at a sufficiently low liquid velocity and in the region of small dimensions. Conversely, when the Peclet number is large, the

concentration distribution is mainly by convective transfer and the molecular distribution can be neglected. The diffusion coefficient in liquid is so small that even at low velocities the mass transport by moving liquid begins to predominate over molecular diffusion. The diffusion coefficient may be represented by the Stokes-Einstein equation :

$$D = K' \, T/3e\mu\pi \tag{8}$$

where, K'= Boltzmann's constant,

T = Temperature in degrees absolute,

μ = the dynamic viscosity of fluid, and

e = the particle diameter (Stokes).

It has been shown that diffusion mechanism is significant only for the particles less than 1 μm in diameter.

2.3.2.2 Inertia

Stream lines approaching the grains diverge as the flow passes around it. If the particles have sufficient inertia, they may cross the stream line and collide with the grain. Davis (1952) quantitatively ascertained the effect of impaction using inertial parameter N, given as :

$$N \propto \frac{\rho_p \, d_p^2 \, V_o}{\mu \, d_m} \tag{9}$$

where, ρ_p = density of the particle,

V_o = superficial velocity, and

μ = viscosity.

Ives (1960), Yao (1971), Ison (1969) demonstrated that the importance of inertia is negligible in water filtration because of low velocity of flow, high viscosity and large collector (media) grain diameter but this mechanism assumes significance in air filtration as the velocity of flow is more and air viscosity and diameter of the collector grain is less. A low value of N, indicates that the removal by inertia is insignificant

2.3.2.3 Interception

A particle with a diameter d_p will make contact with the grain wall if the stream line along which a particle travels comes within a distance of $d_p/2$ from the wall. The suspended particles in the flow may touch the grain and get removed from the flow. Stein (1940) introduced this well-known concept of air-filtration to water filtration. Ison and Ives (1969) & Yao (1971) investigated this experimentally and characterized its effect as :

$$I \propto (d_p)/(d_m) \qquad (10)$$

2.3.2.4 Gravity Settling

Consideration of each pore as a micro-settling basin with low loading rates is a logical extension of pre-treatment mechanism. Settling velocities or terminal velocities are calculated based on Stokes Law :

$$V_s = \frac{g d_p^2 \,(\rho_p - \rho_l)}{18\mu} \tag{11}$$

where, V_s = settling velocity, cm/sec.,

ρ_p = density of the particle, gm/cc, and

ρ_l = density of the liquid, gm/cc.

Ranz (1951) proposed the removal efficiency (η) due to gravity settling by an individual media grain is the ratio of the settling velocity of the particles to the superficial velocity of fluid (V_o).

$$\therefore \quad \eta = \frac{g}{18} \; \frac{(\rho_p - \rho_l)}{} \; \frac{d_p^2}{V_o} \tag{12}$$

$$\text{Therefore } (\lambda_o)GS = \frac{1.5\,(1-f_o)}{d_m}\,\eta$$

$$= \frac{g}{12} \; \frac{(1-f_o)\,(\rho_p - \rho_l)}{\mu\, V_o} \; \frac{d_p^2}{d_m} \tag{13}$$

where, (λ_0) GS = clean bed filter coefficient due to gravity settling.

As the equation is derived for the horizontal flow, Agrawal (1966 questioned the applicability to the down flow filters. Ives (1971) supported it by asserting non-vertical flow through bed capillaries and presence of low velocity zones favouring settling according to Stoke's Law. Gravity settling is a negligible transport mechanism for particles less than 1 μm in diameter regardless of specific gravity but the particles greater than 25 μm will always be affected.

The variation in interstitial velocities about mean velocity will greatly influence this mechanism, especially at the end of the run when many low velocity micro basins have become clogged and the beginning of the run when these have the greatest capacity.

2.3.2.5 Hydro-dynamic Transport

According to Ives (1971), a shear field exist in a laminar flow pattern. Spherical particles travelling within this field experience rotation due to the velocity gradients, in turn spherical particles create a spherical flow field and cause the particles to migrate across the shear field. As the particles are not spherical and deformable and due to non-uniform shear field, the particles will be

deflected in an irregular, impredictable way. As a result, the particles will exhibit an apparently random drifting motion across the stream lines which may cause collision with the grain surface. The complexity of filter pore geometry defies theoretical analysis in fluid shear fields.

2.3.3 Attachment Mechanism

Once the particle has been transported within a close proximity of the filter grain surface, the other mechanisms come into play to cause the adherance of the particle on to the grain and true removal from the fluid. The attachment to great extent determines the filter efficiency and is controlled by surface forces related to the particles and the filter grains. According to Ives (1971), electrical double layer interactions, molecular forces and mutual adsorption are the main attachment forces; however, because of low magnitudes of the particle and filter grain surface potentials and their extremely small range of action in natural waters, electrical effects are negligible in most cases. O'Melia & Stumm (1967) have indicated that electrostatic and chemical forces and the reaction at the surfaces are very active in attachment.

2.3.3.1 Molecular Forces

Van der Waal's are the cause for the physical

adsorption of molecules. Van der Waals forces between two molecules is the molecular attraction force which varies inversely as the sixth power of the separation distance. Shaw (1967) showed the attractive forces between the two spheres :

$$F = (\pi n^2/12r^2)\ (d_1 d_2/(d_1+d_2)) = \frac{A}{12r^2}\ \frac{d_1\ d_2}{(d_1+d_2)} \qquad (14)$$

where, F = attractive force,

n = number, atoms/cm^3,

r = distance between the two spheres,

d_1 and d_2 = the diameter of the two spheres, cms, and

A = Hamaker constant.

This equation is applicable to a limiting distance of 100 A°. Mackrle and Mackrle (1961) proposed that the particles entering the adhesion space were assumed to be removed and the motion of the particles in the adhesion space was governed by the Van der Waal's forces.

2.3.3.2 Mutual Adsorption

This mechanism is due to the bridge formation by polymers such as polyelectrolyte or hydrolysis products of aluminium or ferric salt by having one end attached to grain surface and the other to the particle (Ives, 1966).

For the effective attachment the length of the polymer chain must be at least greater than the range of electrical repulsion.

2.3.3.3 Electrokinetic Phenomena

According to Agrawal (1966), the eletrokinetic phenomena by themselves constitute a complete removal mechanism (transport and the attachment) as these mechanisms considers both physical and chemical parameters of the system. The electrical double layer is formed when a liquid in contact with the solid develops a charge on the solid surface, usually due to adsorption of ions from solution and this solid surface is surrounded by ions of opposite charge, known as counter ions. Thus, an electrical double layer is formed and there will be an establishment of potential difference (streaming potential) when the liquid is passed through a porous media which results in shearing of electrical double layer. The expression for streaming potential is :

$$\Delta E = \frac{\zeta \, \Delta H D_1}{C_1 K' \pi \mu} \qquad (15)$$

where, ΔE = streaming potential, mV,

ζ = Zeta potential, volts,

ΔH = pressure drop across the filter, cm,

K' = electrical conductivity, mho/cm.,

μ = dynamic viscosity, poise, and

D_1 = dielectric constant .

Agrawal (1966) developed the following equation to relate the effect of electrokinetic forces to the filter coefficient when electro-kinetic mechanism is dominant in the removal of particles.

$$\lambda_o = \frac{7\ (1-f_o)}{d_m} E' \qquad (16)$$

and

$$E' = \left(\frac{K_2 \zeta_p^2 N_1 d_m}{\mu V_o} - \frac{K_1 \zeta_m \zeta_p d_p}{\mu V_o} \right) \qquad (17)$$

where, λ_o = clean bed filter coefficient,

f_o = clean bed porosity,

d_m = diameter of filter media, cm.,

d_p = diameter of the suspended particles, cm.,

ζ_m = zeta potential of the media,

ζ_p = zeta potential of the particle,

N_1 = concentration of the particles, nos./ml.,

μ = viscosity, poise,

V_o = superficial velocity of filtration, cm/sec., and

K_1 and K_2 = constants.

Electrokinetic phenomena depends on the type of suspension and media surface charges and there is a possibility of improvement in filtration efficiency by using new suitable types of media.

2.3.4 Overall Filter Coefficient Due To Different Mechanisms

The various mechanisms proposed for explaining the filtration phenomena indicate quite different relationship between the filter coefficients and the major physical parameters like d_m, d_p, μ, V_o. Conditions for predominance of a particular mechanism have also been investigated by many workers but from practical point of view, a simpler method for evaluating the overall filter coefficient is required. Again, contributions by various mechanisms cannot be strictly additive. Yao et al. (1971) introduced a collision efficiency factor to combine the effects due to attachment and transport mechanisms.

$$\frac{\partial C}{\partial L} = - \frac{1.5\ (1-f_o)}{d_m} \alpha \eta\ C \tag{18}$$

The single collector efficiency, η is assumed to be an additive effect of the transport processes whereas α is the collision efficiency factor to represent the overall effect of the attachment mechanisms. For completely destabilized particles the collision efficiency factor $\alpha = 1$; η depends not only upon the filtration velocity, media size, and water temperature but also in a significant manner on the density and size of the particles to be filtered.

2.3.5 Effect Of Clogging On Filter Coefficient

According to Camp (1964), the impediment modulus (λ) is a function of not only the specific deposit, σ, but also of sand size, porosity, the floc size, the adhesive characteristics of the floc and the effect of viscous shearing forces in retarding the floc removal. Also, the deposition of suspended solids affect the filtration efficiency. Camp (1964) has also proposed that λ is approximately proportional to the square of the mean velocity gradient in pores and suggested that adhesive characteristics of floc, or sand or both are less at lower depths. Iwasaki (1937) proposed that owing to the action of gravity the diverting particles from the flow streamlines would be localized to form domes in the filter grain surfaces and due to increase in surface area for deposition and from geometrical considerations λ increases linearly with deposit :

$$\lambda = \lambda_0 + c\sigma \qquad (19)$$

where, λ_0 = initial filter coefficient,

c = system constant, and

σ = specific deposit or floc volume concentration

Ives (1960) proposed that increasing deposition would cause the filter pores to become gradually constricted resulting in

straightening of the flow stream lines, increase in interstitial velocity and reduction in surface area available for deposition and so filter coefficients would decrease at the end of the run. He proposed the modified equation :

$$\lambda = \lambda_o + c\sigma - \frac{\phi_1 \sigma^2}{f_o - \sigma} \tag{20}$$

where, σ = specific deposit, and

c and ϕ_1 = system constants .

Ives and Sholji (1965) related the constants λ_o, c and ϕ_1 with three main physical parameters of filtration :

$$\lambda_o = \frac{K_1}{d_m \cdot V_o \mu^2} \tag{21}$$

$$c = \frac{K_2}{d_m \cdot V_o \mu^{1.2}} \tag{22}$$

$$\phi_1 = \frac{K_3}{d_m \cdot V_o \mu^2} \tag{23}$$

where, K_1, K_2 and K_3 are constants depending upon the properties of media and suspension. For determining the constants a pilot plant study would be necessary. The relationship proposed are based on a laboratory study of one filter system and hence data from field studies are required

for more general application. Fox and Cleasby (1966) observed that for the filtration of hydrons ferric flocs, Ives relationship did not hold good. O'Melia and Stumm (1967) reported that dependence of λ_o on d_m and V_o would vary depending on the dominant mechanism(s) operative in the filter and hence a method for estimating the values of K_1, K_2 and K_3 is necessary in a particular filter system within a practical accuracy limit. Ives (1967) considered all the effects due to deposition like increase in grain diameter, decrease in capillary rise and increase in interstitial velocity together and expressed a relationship :

$$\lambda = \lambda_o \left(1 + \frac{\sigma p}{f_o}\right)^x \left(1 - \frac{\sigma}{f_o}\right)^y \left(1 - \frac{\sigma}{\sigma_u}\right) \qquad (24)$$

where, p = constant depending upon the grain size,

σ_u = ultimate specific deposit which causes the velocity to rise to the limiting, inhibiting velocity, and

x and y = emperical constants.

According to Mohanka (1969), in case of multilayer filtration the term $\left(1 - \frac{\sigma}{\sigma_u} \right)$ should be raised by some power and the expression for becomes :

$$\lambda = \lambda_o \left(1 + \frac{\sigma p}{f_o} \right)^x \left(1 - \frac{\sigma}{f_o}\right)^y \left(1 - \frac{\sigma}{\sigma_u}\right)^z \qquad (25)$$

where, z is a function of filter media and filtering velocity.

It is interesting to note that λ_0 is a multiplying factor in the new equation instead of an additional term in the previously proposed semi-emperical equation. So the change in filter coefficient also depends on the initial coefficient.

2.4 Hydraulics Of Filtration

2.4.1 Kozney-Carman Equation For Flow Through Porous Media

Darcy's fundamental equation of permeability based upon the measurement of flow of water through sand and sand stones may be represented as :

$$V = K'' \frac{\Delta H}{L} \tag{26}$$

where, K'' = coefficient of permeability,

ΔH = pressure difference,

L = depth or thickness of the bed, and

V = rate of flow of water across a unit cube of sand at unit pressure head.

The equation is closely analogous to Poiseuille's Law for the flow of viscous fluid through a circular capillary;

$$V = \frac{d_e^2}{32\nu} \Delta H.g/L \qquad (26a)$$

where, d_e = diameter of equivalent channel,

ν = kinematic viscosity of fluid, poise.

First extension of simple D'arcy Law was made by Dupit who proposed that the apparent velocity V must be less than the actual velocity in the pores and so the true velocity of flow at the pores must be $\frac{V}{f}$ and hence :

$$V = f\ K'' \frac{\Delta H}{L} \qquad (27)$$

According to Carman (1937), Schiller showed that the flow in a pipe of non-circular section, to an acceptable degree of accuracy could be co-rrelated with that of circular pipes if the pipe diameter was replaced by the hydraulic radius m where :

$$m = \frac{\text{volume of fluid in pipe}}{\text{surface presented to fluid}}$$

So, if this expression is applied to a granular bed, $m = \frac{f}{S}$, where f is the porosity and S is the specific surface area.

For viscous flow, the method of Blake (1922) gives rise to the following form of D'arcy equation :

$$\frac{\Delta H.g.f^3}{L\ V\ \nu\ S^2} = K''$$

$$V = \frac{f^3}{K'' \nu S^2} \quad \frac{\Delta H . g}{L} \tag{28}$$

According to Carman (1937), Kozney derived this equation assuming that the granular bed is a group of parallel, similar (capillary) channels such that the total internal surface and the total internal volume are equal to the particle surface and to the pore volume respectively, in the bed itself so that the value of $m = \frac{f}{S(1-f)}$. Further, he mentioned that owing to the tortuous character of flow through a granular bed, the length of equivalent channel should be L_e where L_e is greater than the total depth of the bed L. The general equation of a stream line flow through a channel :

$$V_o = \frac{m^2}{K_o . \nu} \quad \frac{\Delta H \ g}{L_e} \tag{29}$$

where K_o depends upon the shape of the cross-section of the channel. Since the flowing liquid has to follow the sinuous path through a porous media, it covers a greater distance than the actual depth of the bed. There will be an increase of pore velocity by the ratio (L_e/L) and decrease of $\Delta H/L$ by the same ratio. Therefore,

$$V = \frac{f \ m^2}{K_o . \nu} \quad \frac{\Delta H . g}{L} \left(\frac{L}{L_e}\right)^2 \tag{30}$$

$$= \frac{f \cdot f^2}{K_o \cdot \nu \, S^2 (1-f)^2} \quad \frac{\Delta H \cdot g}{L} \; (L/L_e)^2 \qquad (30a)$$

$$\frac{\Delta H}{L} = K_o \left(\frac{L_e}{L}\right)^2 \frac{S^2 \nu}{g} \cdot V \cdot \frac{(1-f)^2}{f^3} \qquad (31)$$

In this expression the incalculable parameter is $K_o \left(\frac{L_e}{L}\right)^2 = K$, known as Kozney-Carman constant. Carman (1937) experimentally showed that for wide range of geometrical shapes, the variation in K_o was between 1.52 to 3.0 with an average of 2 to 2.5. He suggested a value of 2.5 for K_o and 2.0 for $\left(\frac{L_e}{L}\right)^2$ for consolidated porous media and this leads to a value of 5.0 for Kozney-Carman constant, K.

There are many variables in the Kozney-Carman equation as their values change with the progress in filtration. The first is the change in porosity due to clogging. The second is the change in surface area of the media grain due to deposition. This surface area change depends upon the amount and the mode of deposition and the characteristics of flow. The third variable is the change in the tortuosity factor $\left(\frac{L_e}{L}\right)^2$ due to deposition. The length of stream line flow in a clear filter which adheres to the surface of the grain, the length of the stream line has to traverse moving from one grain to another is $\pi D/2$. But, after deposition if the stream line is straight across the grain,

then this length is D, where D is the diameter of the grain. So, the change in $(\frac{L_e}{L})^2$ will vary from 1 to 2.5. The fourth variable is the change in Carman shape factor, K_o. This will change due to change in the shape of the cross-section.

Fair and Hatch modified the original form of the Kozney-Carman equation (Eq. 31) in order to make it more applicable for prediction of filter headloss. Assuming $S = \frac{6}{\phi\, d_m}$ for filter media grain diameter d_m with sphericity ϕ, Eq. 31 reduces to (Hudson, 1969) :

$$\frac{\Delta H}{L} = \frac{K_o\,(\frac{L_e}{L})^2 \times 36}{\phi^2}\ \frac{\nu}{g}\ \frac{(1-f)^2}{f^3}\ \frac{V}{d_m^2}$$

$$= JS_o^2 \cdot \frac{\nu}{g}\ \frac{(1-f)^2}{f^3}\ \frac{V}{d_m^2} \qquad (31a)$$

2.4.2 Headloss Equations in Filtration

Kozney-Carman equation or its modification by Fair and Hatch is not applicable for estimating the headloss with progress in filtration. This is because it is difficult to predict from theoretical or intuitive considerations the change in the various parameters with filtration time. As a result of this many investigators have proposed different equations for predicting filter headloss during filtration.

Deb (1969) proposed a model in which he assumed that the deposition took place uniformly on the grain surfaces. He considered the point of contact between the grains and their effect on specific surface and the change in porosity, surface area and Kozney-Carman constant K due to deposition in his model. The change in the value of K and specific surface area with specific deposit σ was determined experimentally and used to determine headloss at any time during filtration :

$$\frac{H}{H_o} = \left(1 + G\left(1-10^{-K'''\sigma}\right)\right) \left(\frac{f}{f-\sigma}\right)^3 \qquad (32)$$

where, G and K''' are emperical constants and the values are 3.2 and 13.3 respectively.

Ives (1960) in his model introduced one constant for the change in surface area and Kozney-Carman constant during filtration :

$$\frac{H}{H_o} = \left(\frac{K^1}{K}\right) \left(\frac{r_1}{r_o}\right) \frac{(1-f+\sigma)^2 \cdot f^3}{(f-\sigma)^3 (1-f)^2} \qquad (33)$$

where, $r_1 = \dfrac{\text{area}}{\text{volume ratio of coated filter grains}}$

Mackrle (1965) suggested a mathematical model expressing the change in specific surface with progress in filtration :

$$\frac{H}{H_o} = \left(1 + \frac{p\sigma}{f}\right)^3 \left(1 - \frac{\sigma}{f}\right)^{-3/2} \tag{34}$$

Using Mackrle's mathematical model and Horners (1968) method, Mohanka (1969) proposed an equation for headloss during filtration :

$$\frac{H}{H_o} = \left(1 + \frac{p\sigma}{f}\right)^2 \left(1 - \frac{\sigma}{f}\right)^{-1} \tag{35}$$

where, $p = \dfrac{29}{S^{0.65}} = 9.05 \left(\dfrac{\phi\, d_m}{1-f}\right)^{0.65}$,

ϕ = sphericity of the filter grain, and

d_m = average media grain size, cm.

In his proposed headloss equation, Mohanka assumed that Kozney-Carman constant remained unchanged during filtration and flow remained laminar throughout. His equation takes into account the change in surface area and porosity of the bed.

Camp (1964) assumed that the clogging formed a sheath on the surface of the grains and he did not consider the point of contact between the grains in the filter. The Camp's equation for headloss during filtration does not consider the change in Kozney-Carman constant, K during filtration :

$$\frac{H}{H_o} = (1 + \frac{\sigma}{1-f})^{4/3} \quad (1 - \frac{\sigma}{f})^{-3} \qquad (36)$$

Shektman (1961) considered change in porosity in his model while other variables remained constant during the filtration :

$$\frac{H}{H_o} = (1 - \sqrt{\frac{\sigma}{f}})^{-3} \qquad (37)$$

Sakthivadivel (1972) considered the change in porosity during filtration and introduced one constant to take into account other changes like specific surface area & Kozney-Carman constant :

$$\frac{H}{H_o} = \frac{(1-f+\sigma)^2}{(f-\sigma)^3} \quad \frac{f^3}{(1-f)^2} \cdot \frac{1}{\xi^2} \qquad (38)$$

Deb (1969) determined experimentally the change in surface area due to deposition with the change in Kozney-Carman constant and plotted

$$(\frac{K'_o}{K}) \; (\frac{L'_e}{L}) \; (\frac{L}{L_e})^2 \; (\frac{S'}{S})^2 = \xi^2$$ as a function

of specific deposit, σ. The graph indicated that for $\sigma > 0.06$, ξ^2 remained almost constant and equal to 4.0. He also found out the variables in the filters with other unknowns and based upon this he proposed his headloss

equations whereas the others incorporated the change of surface area and the porosity in the proposed model itself. According to Sakthivadivel (1972), Camp's equation gives the lowest headloss upto $\sigma = 0.17$, then the values are greater than those predicted by Mohanka's model. Deb's model gives fairly high values of $(\frac{H}{H_o})$ compared to the other models. Ives model cannot be compared with the others because he did not determine the constant experimentally. According to Sakthivadivel (1972), Shektman in his equation took into account only porosity and assumed that the other variables remain constant during filtration and equal to 1. Mackrle (1961) in his model considered the change in surface area and porosity but has not considered the change in tortuosity and Carman shape factor.

It is clear that during filtration the porosity, specific surface area, tortuosity and Kozney-Carman constant changes differently and these changes also depend upon the type of media and suspension. There is no method presently available to determine the change in tortuosity and Kozney-Carman constant. Again, it is not known whether the flow is laminar or turbulent with increase in specific deposit during filtration.

2.5 Present Methodology For Determination Of Specific Deposit

The mathematical models proposed by various investigators to predict headloss during filtration clearly indicate the significance of the specific deposit term and hence its estimation during filtration would be meaningful. Consequently, several investigators attempted experimental estimation of floc volume concentration or specific deposit in a filter.

2.5.1 Centrifuging

Baylis (1931) measured the floc volume in flocculated and settled water by centrifuging. However, according to Mohanka (1969), the floc volume of the influent sample depends upon the speed of the centrifuge. He found that the floc volume of the influent sample was reduced three-folds when the centrifugal force increased from 112 g to 388 g. Hudson (1965) attempted to determine the floc volume concentration from jar test data and proposed that if the time of agitation and the velocity gradient were known the value of ϕv could be calculated from jar test using the equation.

$$\frac{N_t}{N_o} = e^{-\frac{\psi v \, G_1 t}{\pi}} \qquad (39)$$

where, N_o = number of suspended matter originally present,

N_t = remaining free or unflocculated matter after time t,

G_1 = velocity gradient,

v = floc volume concentration, and

ψ = sticking ratio.

The sticking ratio, ψ , depends upon the molecular forces and may be influenced by temperature, pH, zeta potential, particle charges, exchange capacities, coagulant dose and so forth. Hudson (1965) also observed that at any constant velocity gradient below 39, the values of ψv were directly proportional to alum dose. Assuming ψ to be constant, the volume of floc is much greater for low agitation than for high agitation speeds, the difference is as great as 25 folds. Again, the floc formed following rapid mixing appeared to contain approximately $2\frac{1}{2}$ times more solids than that formed when there is no rapid mixing.

2.5.2 In situ Measurement

Ives used radioactively lebelled algae for estimating floc volume concentration by measuring the distribution of algae deposited in the filter with a scintillation counter. According to Ives (1961), Rabinson tried

timing ultrasonic pulse transmission across a clean filter and one clogged with quartz powder but no quantitative results were obtained.

2.5.3 Backwashing

Mohanka (1969) tried several alternatives using iron floc for determination of floc volume concentration. He collected the backwash waste from a filter and the floc was allowed to settle in an Imhoff cone but he did not adopt this method because the results did not fit with his model satisfactorily. Hsiung (1974) found out the average specific deposit from backwash waste by settling the waste in an Imhoff cone. In computing specific deposit values using backwash sludge volume concentration, he assumed that the primary floc occupied the same volume when deposited in the bed as when in suspension and that the agglomerated floc occupies the same volume when retained in the bed as when settled from backwash waste. However, according to Hudson (1969), when a primary floc contacts a filter grain surface the floc structure is likely to alter. Baylis (1931) observed that flocculated matter removed in the filter compacts during the filtration process and the process of compacting causes the removed material to adhere more firmly to filter media or other compacted material. The

compaction of the flocculated matter in the filter backwash water is much more than the influent suspension applied to the filters. Sriramulu et. al. (1975) questioned Hsiung's assumptions and proposed that the alteration of the floc structure may be due to the expulsion of a part of intra-floc water, but due to agglomeration in the bed, interfloc water would build up again. Further, the agglomerated floc deposited in the bed is sheared off during backwashing and it is likely that a part of the interfloc water will be lost. Due to agglomertion in the Imhoff cone during subsequent settling the interfloc water will again build up but may be compared to a different extent to that in the floc deposited in the bed. Concievely, the actual floc volume concentration or specific deposit in a filter bed may not be same as estimated by settling the backwash waste in an Imhoff cone. Nevertheless, estimation of specific deposit using backwash waste or even primary floc definitely appears attractive as well as practically useful and hence merits more careful examination from both experimental and theoretical viewpoints.

3. SCOPE OF THE INVESTIGATION

From the preceeding discussion it is apparent that the physicochemical phenomena involved in filtration of water through granular media are rather complex and more understanding is required for efficient design and effective operational control of water filtration plants. In order to be able to predict the performance of a filter employing a particular media-suspension combination, a background knowledge of the specfic deposit would be practically meaningful. This would allow the prediction of filter performance in terms of headloss development using the presently available mathematical models. Considering the work already conducted in this regard, it appears appropriate to undertake thorough investigation of specific deposit change in a filter with time and depth including its computation using a simple, acceptable laboratory technique which might as well lead to its estimation using settling test data of the primary suspension or the backwash waste.

The present investigation was initiated to study the nature of specific deposit change with filtration time and filter depth using several media suspension combination, e.g., sand and coal, and kaolinite turbidity alum coagulated kaolinite turbidity and wastewater. It was thought that this

would also lead to a possible generalization of specific deposit variation with respect to media-suspension combination. The specific deposit were computed using (i) porosity change depicted by the Fair-Hatch modification of Kozney-Carman equation from the headloss data and the location of the clogging front at a particular filtration time and (ii) settled sludge volume fraction obtained by settling the influent (primary) suspension or backwash waste in an Imhoff cone, volume of water filtered or backwash waste volume, and influent and effluent turbidity (solids concentration) or backwash waste solids concentration. The specific deposit values thus computed were then compared with those obtained using some of the available headloss equations.

4. LABORATORY SETUP AND EXPERIMENTAL PROCEDURE

4.1 Filtration Setup

The filtration setup consisted of a glass filter column (corning glass), 2.9 cm ID and 1.0 m long, with suitable ports for headloss measurements (Fig. 1). A 60 l plastic container equipped with a stirrer served as the influent storage tank and provided a filtration head of 250 cm. The filtration rate was monitored through a rotameter and kept constant during an experiment by manually adjusting the filter outlet.

4.2 Filter Media

The filter media used were Jamuna sand and Giridih bituminous coal of geometric mean size 0.505 and 0.845 mm, respectively. The sand and coal were washed several times in tap water, dried at 103 °C for 24 hr. before using.

The media characteristics were : Specific gravity 2.6 (sand) and 1.4 (coal), porosity in filter column 0.51 (sand) and 0.567 (coal), sphericity 0.6 (sand) and 0.547 (coal). Specific gravity was determined using a specific gravity bottle.

For determining the porosity in filter column, the following relationship was employed :

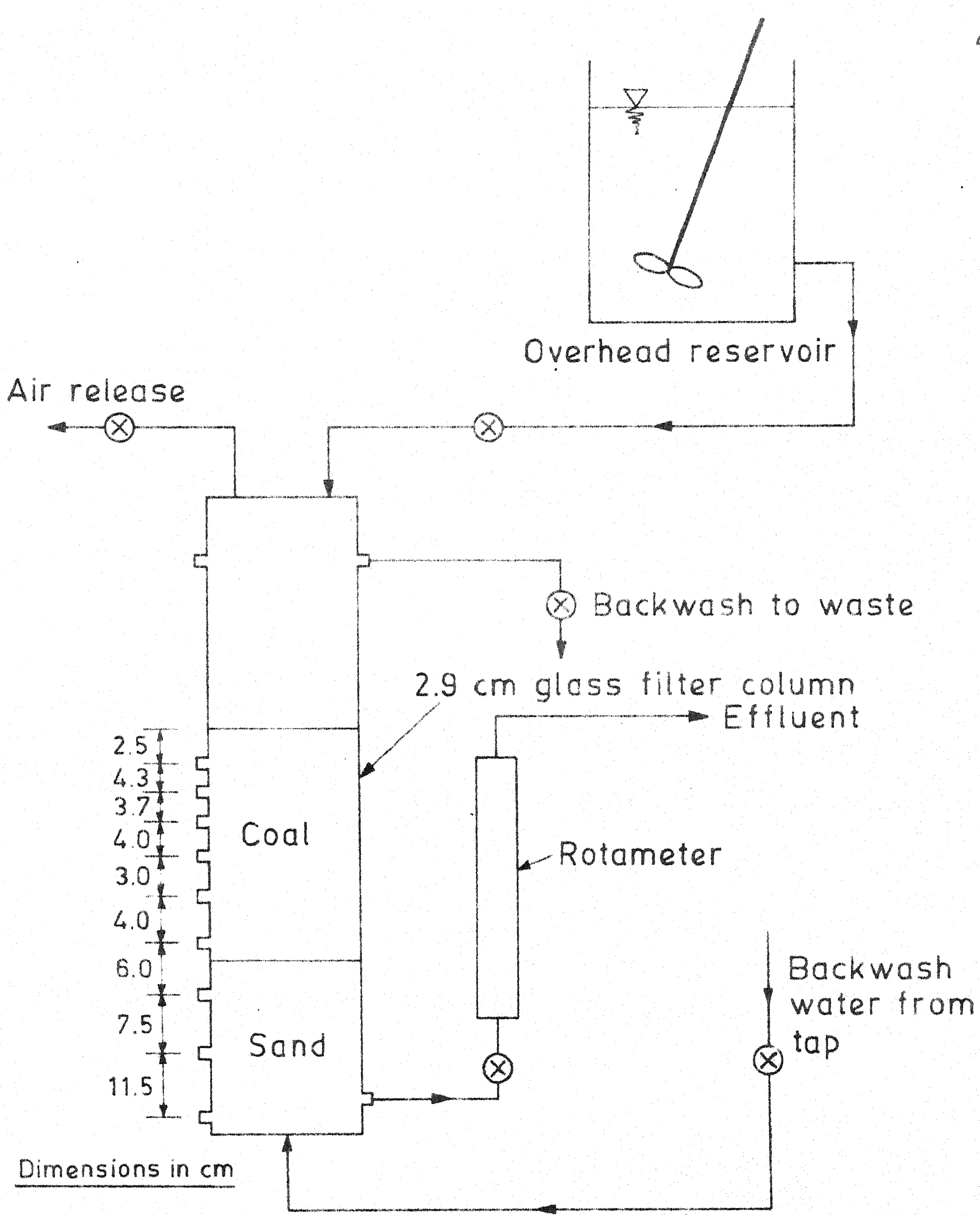


Fig. 1. Schematic flow diagram.

$$\text{porosity (f)} = \frac{V_t - \frac{W_s}{\gamma_s}}{V_t} \tag{40}$$

where, V_t = volume of the media in filter column, cm^3

W_s = weight of the media, gm, and

γ_s = specific gravity of the media.

Sphericity of the media was determined by measuring the discreate particle settling velocity and employing the relationship.

$$\text{Sphericity } (\emptyset) = \frac{V_s}{\frac{g}{18} \frac{(\rho_s - \rho_1)}{\mu} d_m^{\ 2}} \tag{41}$$

where, V_s = stokes settling velocity of the equivalent volume sphere, cm/sec.

4.3 Suspension and Preparation

4.3.1 Kaolinite Suspension

Kaolinite suspension was prepared by vigorously stirring a desired quantity of 200 mesh Kaolinite clay in tap water and transferring the requisite quantity of supernatant (2 hr settling) to the storage tank containing about 60 l of tap water to produce a desired influent turbidity which ranged between 240 and 330 NTU.

4.3.2 Coagulated Kaolinite Suspension

The kaolinite suspension was prepared as before, coagulated with commercial alum using slightly less than optimum jar test dose and the supernatant following 30 min settling was used. The influent turbidity ranged between 45 and 88 NTU.

4.3.3 Wastewater

I.I.T. Kanpur campus wastewater settled for 6 hr. followed by heating to 80°C to avoid fouling and subsequent cooling to room temperature was used as the influent suspension. The turbidity of this suspension ranged between 18 and 30 NTU.

Standard curves relating turbidity and suspended solids concentration were prepared for all the four suspensions (Fig. 2-2a). A Hach Model 2100A turbidimeter was used for turbidity measurement, Suspended solids concentrations were determined by filtering a known volume of the suspension through a 0.45 μm membrane filter.

4.4 Filter Run And Specific Deposit Determination

Filtration experiments were conducted employing sand (46.5 cm), coal (46.50 cm) or coal-sand (23.25 and 23.25 cm) as filter media for the three influent suspensions,

viz., kaolinite and coagulated kaolinite suspensions, and wastewater. For destablized kaolinite suspension, one experiment was performed with coal-sand dual-media. Normal rapid sand filtration rates, viz., 81 and 162 lpm/sq.m. (2 and 4 gpm/ft^2), were used. The influent suspension in the overhead storage tank was kept stirred continuously to minimize variation in influent turbidity. Measurements of head loss and turbidity (influent and effluent) were done at 30 min intervals. Duration of a filter run varied between 6 and 12 hr. depending on media-suspension combinations. At the end of a filter run the filter column was backwashed (50% bed expansion) with tap water and the entire backwash waste was collected, its volume measured and turbidity determined.

For <u>in situ</u> specific deposit determination, porosity change of the filter bed with filtration depicted by Fair and Hatch (1933) modification of Kozney-Carman equation (Eq.31a) was used :

$$i = \frac{j\, S_o^2 \cdot \nu}{g} \quad \frac{(1-f_o)^2}{f_o^3} \quad \frac{V}{dm^2} \qquad (31a)$$

where where, i = hydraulic gradient,

jS_o^2 = filter media geometry characteristics,

ν = kinematic viscosity,

f_o = clean bed porosity,

g = acceleration due to gravity,

V = approach velocity, and,

d_m = filter media grain diameter

For a filter bed of stafed media size and for a fixed filtration rate (Hudson, 1969)

$$\frac{i_t}{i_o} = K \frac{(1-f_t)^2}{f_t^{\,3}} \tag{42}$$

where, i_t = hydraulic gradient at time t,

i_o = initial hydraulic gradient, and,

K = proportionality constant

$= \dfrac{f_o^{\,3}}{(1-f_o)^2}$ where, f_o = clean filter bed porosity

The emplicit assumption in Eq. 42 is that changes in the terms jS_o^2 and $d_m^{\,2}$ with filtration are negligible. According to Camp (1964), the value of jS_o^2 increases with increase in Reynolds number in the transition region but remains fairly constant in the laminar flow region which holds good in the

case of a conventional filter as well as the present experimental study. However, since change in term d_m^2 is not considered, any increase in the filter media size due to deposition would lead to an estimation of f_t slightly higher than the actual value. Knowing the filter bed porosity (f_o) and the hydraulic gradient values, change in filter bed porosity at time t was computed from the above equation. Hence, the volume of floc deposited in the bed :

$$(f_t - f_o) \times \text{(filter bed volume from top upto the depth of floc penetration)}$$

The depth of floc penetration (clogging front distance from top of filter bed) (Advin, A. and Rebhun, M. 1974) obtained from a plot of headloss Vs. depth of filter bed with time (Hudson, 1969). Therefore, average specific deposit :

$$\sigma = \frac{(f_t - f_o) \times \text{(filter bed volume from top upto the depth of floc penetration)}}{\text{total filter bed volume}} \qquad (43)$$

Inability to take into account the change in filter media size due to deposition would possibly lead to an estimation of *in situ* specific deposit value slightly lower than the actual value.

Specific deposit was also computed by settling (12 hr)

the primary suspension and backwash waste in Imhoff cones (Hsiung, 1974) :

$$\sigma = \frac{Q\ (C_o - C)}{C_s \cdot L}\ t \tag{44}$$

where, σ = average specific deposit in total depth L, volume/volume,

Q = filtration rate,

C_o = influent suspended solids concentration, mg/l,

C = effluent suspended solids concentration, mg/l,

C_s = sludge volume concentration, mg/l,

= (suspended solids concentration of the suspension, mg/l) $\div$ (Imhoff cone settled sludge volume fraction, volume/volume), and

L = filter bed depth.

5. RESULTS AND DISCUSSION

Summarized data of typical experiments are presented here in tabular or graphical form followed by discussion of the results. Many experiments were repeated to check reproducibility.

In all twelve filtration experiments (filter runs) were performed using various media-suspension combinations. A summary of the experimental conditions are shown in Table 1. Fig. 3 is a typical plot showing the headloss, hydraulic gradient and floc penetration for the sand-kaolinite filter runs. Similar plots for the remaining runs are included in the Appendix (Fig. A1 - A11). Location of the clogging front from the top of the filter bed with respect to time is obtained by joining the points of inflection of the headloss Vs. depth of bed curves. Floc penetration Vs. time shows the advancement of the clogging front in the filter bed.

In situ specific deposit against time and depth is computed by knowing the change in the filter bed porosity using the Fair-Hach modification of the Kozney-Carman equation and the hydraulic gradient as well as the depth of floc penetration. Equation 43 is used for the computation but the filter bed volume upto the stated depth from the top is considered

TABLE 1

SUMMARIZED EXPERIMENTAL DATA AND SPECIFIC DEPOSIT

Filter Run	Primary (Influent) Suspension	Filter Media	Filtration: Turbidity (NTU) Influent	Filtration: Turbidity (NTU) Effluent	Filtration: Flow Rate ($l/min/m^2$)	Filtration: Time (hr)	Filtration: Headloss (cm) Initial	Filtration: Headloss (cm) Final	Primary Suspension: Suspended Solids (mg/l)	Primary Suspension: Settled Sludge Volume Fraction (vol./vol)	Primary Suspension: Sludge Concentration (mg/l)	Backwash Waste: Volume (l)	Backwash Waste: Suspended Solids (mg/l)	Backwash Waste: Settled Sludge Volume Fraction (vol./vol)	Backwash Waste: Sludge Concentration (mg/l)	Average Specific Deposit (σ): In Situ	Average Specific Deposit (σ): Primary	Average Specific Deposit (σ): Backwash	Average Specific Deposit (σ): Shektman	Average Specific Deposit (σ): Mackrle	Average Specific Deposit (σ): Mohanka
KS1	Kaolinite	Sand	307	36	81	11.08	22.7	62.9	525	0.0022	239×10^3	12.6	855	0.0045	190×10^3	0.053	0.227	0.185	0.044	0.093	0.142
KS1'	Kaolinite	Sand	240	38	162	7.33	51.0	152.0	419	0.0018	349×10^3	12.0	630	0.0031	203×10^3	0.042	0.152	0.121	0.048	0.100	0.153
KS2	Kaolinite	Coal and Sand	330	71	162	11.2	28.0	84.1	560	0.0016	350×10^3	9.25	1440	0.0067	213×10^3	0.065	0.301	0.203	0.050	0.098	0.148
KS3	Kaolinite	Coal	330	100	162	4.67	14.5	18.5	560	0.0016	339×10^3	2.66	658	0.0055	120×10^3	0.01	0.107	0.053	3.45×10^{-3}	0.018	0.0749
CKS1	Coagulated Kaolinite	Sand	65	3.5	81	12.08	21.0	38.0	120	0.0011	105×10^3	4.25	680	0.0042	162×10^3	0.024	0.128	0.058	0.021	0.0597	0.091
CKS1'	Coagulated Kaolinite	Sand	47	2.5	162	9.167	29.8	124.9	100	0.0016	62.5×10^3	5.45	420	0.011	38.2×10^3	0.083	0.338	0.195	0.071	0.1128	0.1747
CKS2	Coagulated Kaolinite	Coal and Sand	45	2.5	162	10.08	26.8	124.1	95	0.00145	65.5×10^3	6.15	370	0.0105	35.2×10^3	0.079	0.291	0.21	0.087	0.1367	0.2104
CKS3	Coagulated Kaolinite	Coal	88	20.0	162	4.5	14.5	45.6	152	0.00425	54.1×10^3	7.40	350	0.0084	41.6×10^3	0.043	0.384	0.203	0.057	0.0915	0.1437
WW1	Wastewater	Sand	24	3.5	81	6.167	22.5	110.0	92	0.25×10^{-3}	384×10^3	2.80	220	0.005	44.0×10^3	0.060	-	0.0457	0.083	0.144	0.2189
WW1'	Wastewater	Sand	19	6.0	162	4.08	49.8	139.2	60	0.2×10^{-3}	300×10^3	3.60	146	0.0025	58.4×10^3	0.048	-	0.0293	0.043	0.094	0.1437
WW2	Wastewater	Coal and Sand	17	6.0	162	6.66	28.1	61.3	52	0.18×10^{-3}	289×10^3	2.80	248	0.0052	47.6×10^3	0.0384	-	0.047	0.028	0.067	0.1022
WW3	Wastewater	Coal	30	15.0	162	6.08	13.5	26.8	128	0.32×10^{-3}	400×10^3	5.30	160	0.0032	50.0×10^3	0.032	-	0.055	0.0273	0.0575	0.0866

Note : Media grain size (d_m), depth (L), porosity in filter column (f) and sphericity (ϕ) were 0.505 mm, 46.5 cm, 0.51 and 0.6 for sand and 0.845 mm, 46.5 cm, 0.567 and 0.547 for coal, respectively.

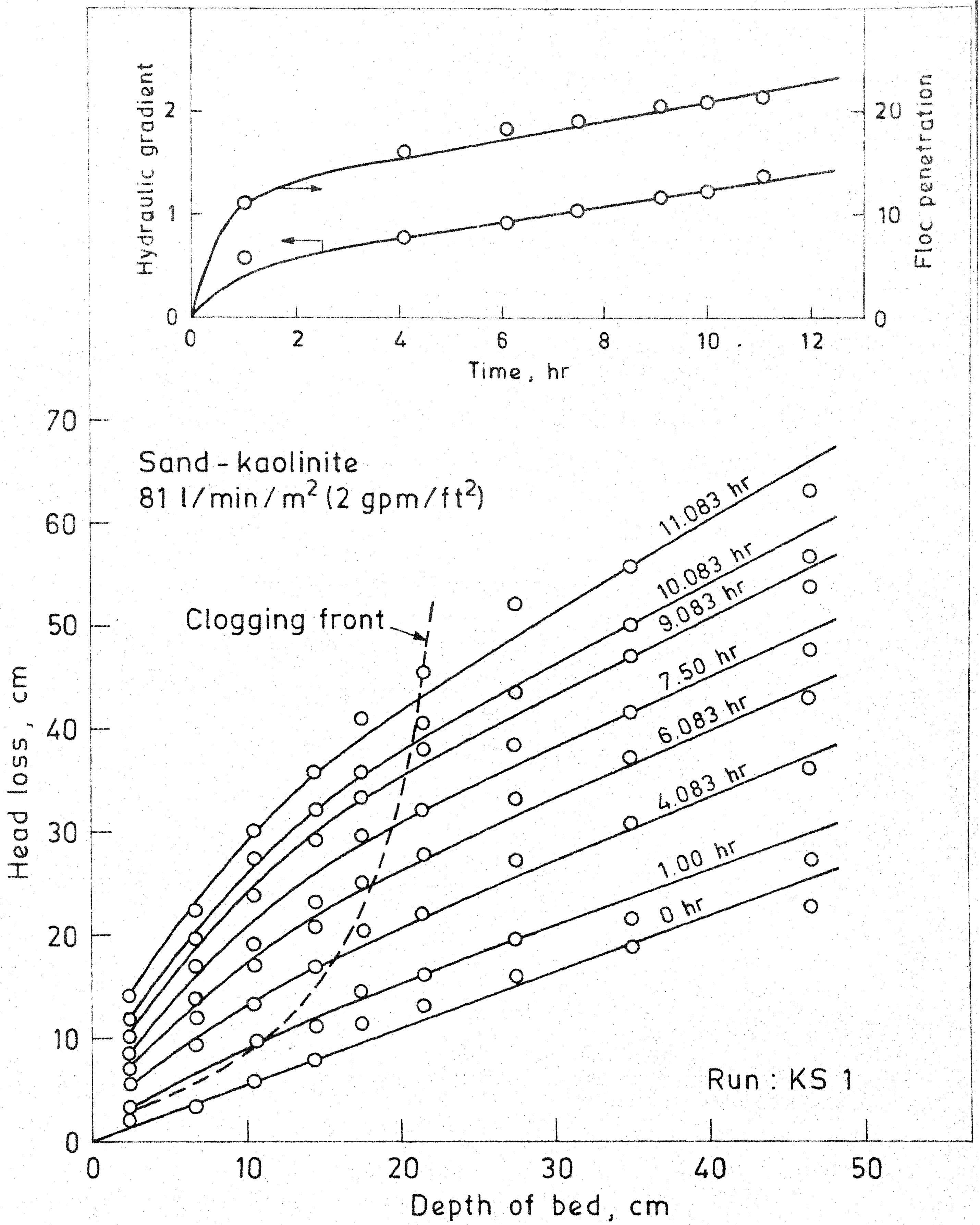


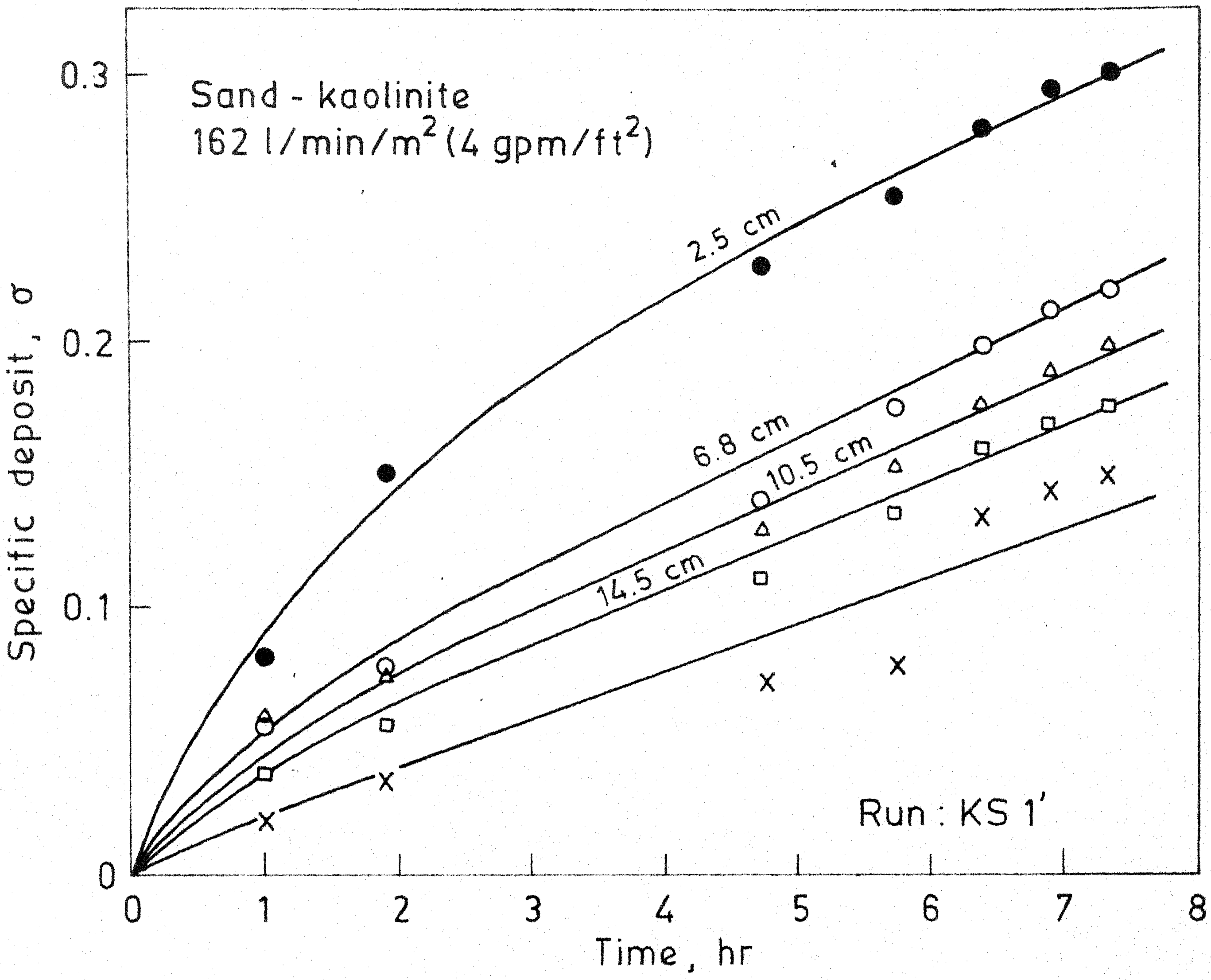
Fig. 3. Head loss, hydraulic gradient and floc penetration in sand - kaolinite system at 81 $l/min/m^2$.

instead of total filter bed volume. Also, in the numerator of Eq. 43 filter bed volume taken is upto the depth of floc penetration or the stated depth whichever is smaller as shown by the floc penetration Vs. time plot. A typical trend of specific deposit change with time and depth is presented in Fig. 4 and such plots for the remaining filter runs are included in the Appendix (Fig. A12 - A16).

The following observations may be generalised from these plots :

(i) Specific deposit varies linearly with the rate of filtration for a specific media-suspension combination. This is also concieveable by considering that specific deposit is directly proportional to the total volume of particulate matter deposited in the filter bed at a given time.

(ii) For a particular media-suspension combination, specific deposit is proportional to the net solids uptake (difference of the average influent and effluent turbidity) by the filter bed. This is also concieveable from the above mentioned consideration. However, this generalisation will be valid as long as there exists a linear relationship between turbidity and suspended solids concentration for a suspension and allows normalisation of specific deposit values for reduced or increased net solids uptake within the linear limit of the turbidity vs. suspended solids plot. It is worthwhile to note here that the above



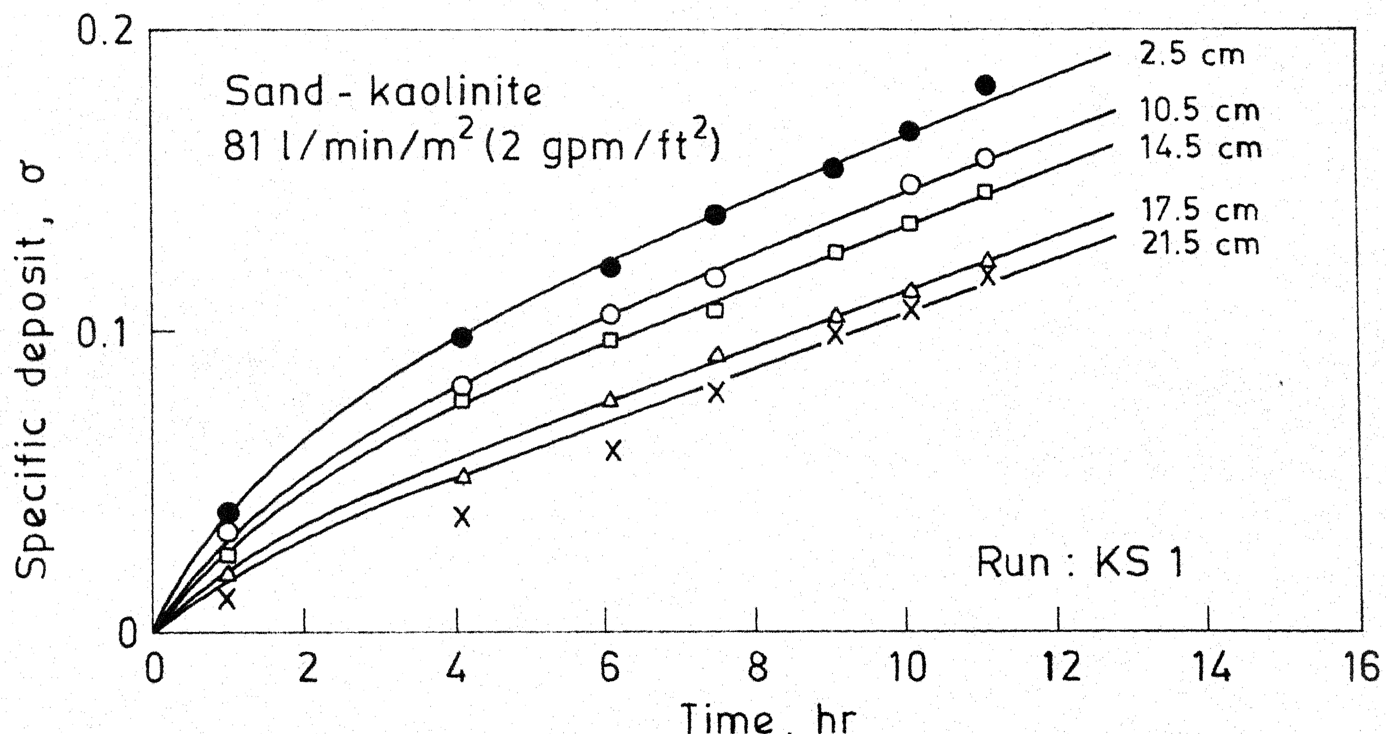


Fig. 4. Specific deposit change during sand filtration of kaolinite suspension.

two generalization hold good provided the clogging front is contained within the filters.

Temporal average specific deposit change (Eq. 43) for various filter media-suspension combination is shown in Fig. 5. However, for the purpose of comparison the average specific deposit values have been normalised for an uptake of 90 mg/l solids concentration (difference between influent and effluent turbidity converted to solids concentration using Fig. 2 and 2 (a)). This normalization is valid on the basis of the foregoing conclusion (ii). It is observed from Fig. 5 that for all the influent (primary) suspensions, the rate of specific deposit build up is faster in the case of sand compared to coal; for coal-sand dual-media this lies somewhere inbetween. Presumably, this is primarily due to finer media grain size in the case of sand. Operational advantage of coal-sand dual-media filter is also indicated here. Regarding the influence of the suspension solids characteristics on specific deposit, it is observed that the rate of specific deposit build up increases in the order kaolinite, kaolinite (coagulated) and wastewater.

This indicates that in the case of organic solids, expulsion of intrafloc water (compaction) is less when it contacts a filter media grain surface or interfloc water buildup is more due agglomeration in the filter bed. Similar

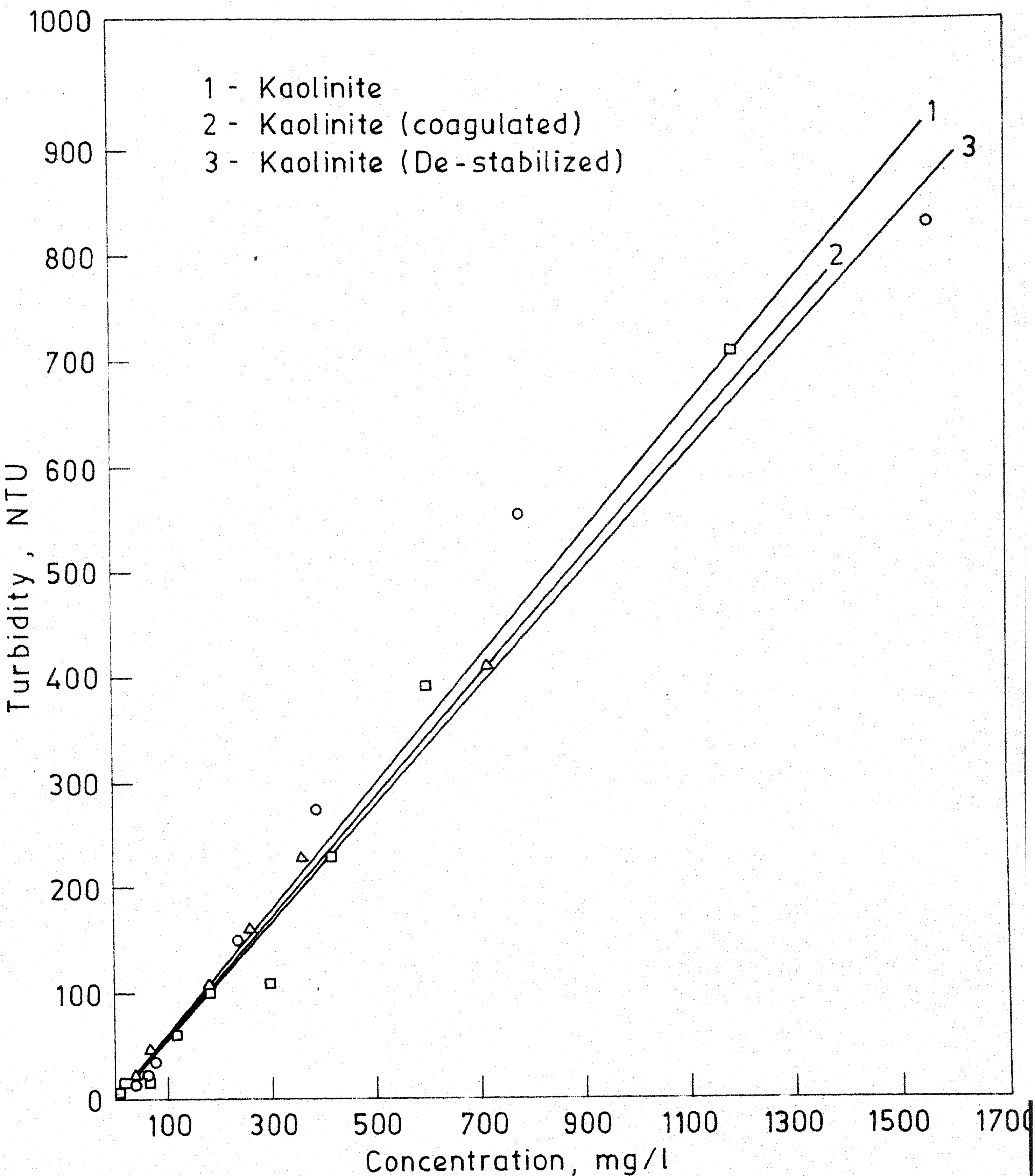


Fig. 2. Turbility vs suspended solids concentration plots for the influent (primary) suspensions.

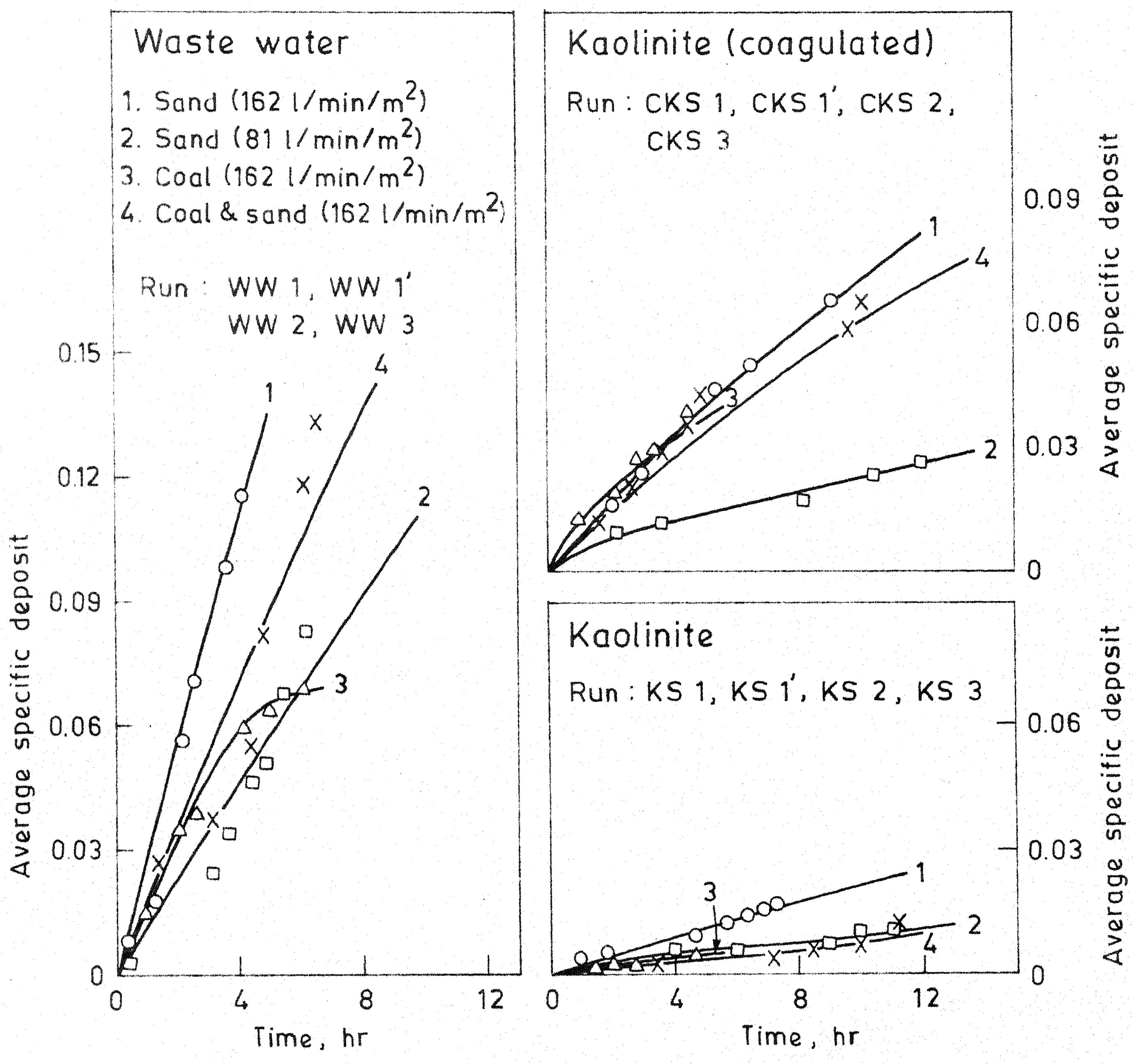


Fig. 5. Temporal average specific deposit change (normalised for 90 mg/l suspended solids uptake) for various filter media - suspension combination.

TABLE 2

RELATIONSHIP BETWEEN IN SITU SPECIFIC DEPOSIT AND SPECIFIC DEPOSIT (PRIMARY) AND SPECIFIC DEPOSIT (BACKWASH)

Primary (influent) Suspension	Filter Media	Specific deposit (primary) / In situ Specific deposit	Specific deposit (backwash) / In situ Specific deposit
Kaolinite	Sand	3.4-4.1	2.9-3.5
Kaolinite	Coal-Sand	4.5	3.1
Kaolinite	Coal	10.6	5.3
Kaolinite (coagulated)	Sand	3.9-5.1	2.4-2.5
Kaolinite (coagulated)	Coal-Sand	2.4	2.7
Kaolinite (coagulated)	Coal	9.0	4.7
Wastewater	Sand	-	0.6-0.8
Wastewater	Coal-Sand	-	1.2
Wastewater	Coal	-	1.8

difference also exists between kaolinite and kaolinite (coagulated) a major fraction of which consists of hydrons aluminium oxide floc.

Table 1 shows the average _in situ_ specific deposit for the twelve filter runs using various filter media suspension combination. Specific deposit values computed from settling of influent (primary) suspension and backwash waste solids, and four headloss equations (Eq. 34-37) are also included. For wastewater, specific deposit (primary) values are not included due to poor settling observed in the Imhoff cone. Table 2 shows the relationship between _in situ_ specific deposit and specific deposit (Backwash) and specific deposit (primary) for the different filter media-suspension combination. Higher specific deposit values obtained through the backwash waste or primary suspension approach in the case of kaolinite and kaolinite (coagulated) indicate that during setting of the backwashed solids interfloc water builds up to a much greater extent compared to that during deposition in the filter bed, and there is a significant degree of compaction when the solids are deposited in the filter bed as observed by Baylis (1931). Hence, the assumptions made by Hsiung (1974) in his backwash approach of specific deposit determination, i.e., that the primary floc occupies the same volume when deposited in the bed as when in suspension and that the agglomerated floc

occupies the same volume when retained in the bed as when settled from the backwash waste, do not hold good. In the case of wastewater solids, the interfloc water build up during settling of the backwashed solids is comparatively much less and probably the agglomerated floc occupies the same volume when retained in the bed as when settled from the backwash waste.

Based on the observations reported here it appears that specific deposit estimation using the backwash approach is more appropriate. For sand or coal-sand dual-media filter receiving alum coagulated influent, specific deposit (backwash) should be divided by a factor of 2.5 to obtain the in situ specific deposit. During filtration of wastewater through sand or coal-sand dual-media filter specific deposit (backwash) is probably identical to the in situ specific deposit.

Several equations are available for predicting filter headloss using filter media characteristics and specific deposit. Four of these equations used by Hsiung (1974), viz., Eq. 34-37, were also employed here to compare the in situ specific deposit values with those obtained by using these equations. This would also indicate the applicability of a particular equation for predicting the filter headloss using the backwash approach of specific deposit estimation. It is

observed from Table 1 that the headloss model proposed by Shektman (Eq. 37) is valid for this purpose because the in situ specific deposit values are comparable to those computed using this model.

To justify the validity of Shektman's model based on comparable specific deposit values obtained using this model and the in situ approach used in the present study, the following discussion may be put forward. The parameters in the Kozney - Carman equation (Eq. 31) that would vary with filtration are porosity (f) tortuosity factor (L_e/L_o), surface area (S), and Carman shape factor (K_o). According to Sakthivadivel (1972), Shektman took explicitly the porosity change as the foremost variable and introduced a constant to take into account the other variables which ultimately was taken as equal to unity. This is also true for the in situ approach used in the present study.

6. SUMMARY AND CONCLUSIONS

An investigation was conducted to study the specific deposit change with filtration time and filter depth using several media-suspension combination, e.g., sand and coal, and kaolinite turbidity, alum coagulated kaolinite turbidity and wastewater. It is observed that for a specific media - suspension combination, specific deposit is proportional to the rate of filtration and the net solids uptake (difference of average influent and effluent turbidity). Specific deposit build up is influenced by both media as well as suspension solids characteristics. It is faster in the case of sand compared to coal and rate of build up increases in the order kaolinite, kaolinite (coagulated), and wastewater.

Based on the observations reported, the backwash approach proposed by Hsiung (1974) appears to be a feasible technique for specific deposit estimation. However, the assumptions made by Hsiung do not hold good under all conditions. It is recommended that for sand or coal-sand dual-media filter receiving alum coagulated influent, specific deposit (backwash) should be divided by a factor of 2.5 to obtain the in situ specific deposit. During filtration of wastewater through sand or coal-sand dual-media filter specific

deposit (backwash) is probably indentical to the *in situ* specific deposit. The equation proposed by Shektman (1961) may be employed thereafter to predict the headloss during filtration.

7. ENGINEERING SIGNIFICANCE AND SUGGESTIONS FOR FUTURE WORK

The findings of the present study add to an understanding of specific deposit change in granular water filter due to solids deposition during filtration. More detailed investigation employing the in situ approach used in the present study may ultimately lead to more rational design of water filtration plants.

The results of this study indicates the feasibility of a practically useful technique for estimating specific deposit which may be employed in predicting the headloss development and efficiency of filtration using some of the available mathematical models.

On the basis of the results obtained in the present study it is felt that further work should be pursued in the following lines :

(i) The relationship proposed between the in situ specific deposit and the specific deposit estimated by the backwash approach should be further checked using a large number of experiments with varying conditions.

(ii) Specific deposit change should be investigated using a number of filter media of comparable grain size and several influent suspensions. This would indicate the effect of the filter media surface characteristics on specific deposit build up, if any.

(iii) A number of influent suspensions with comparable sets of turbidity may be employed to further substantiate the observation that specific deposit build up varies linearly with the net solid uptake.

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APPENDIX - A

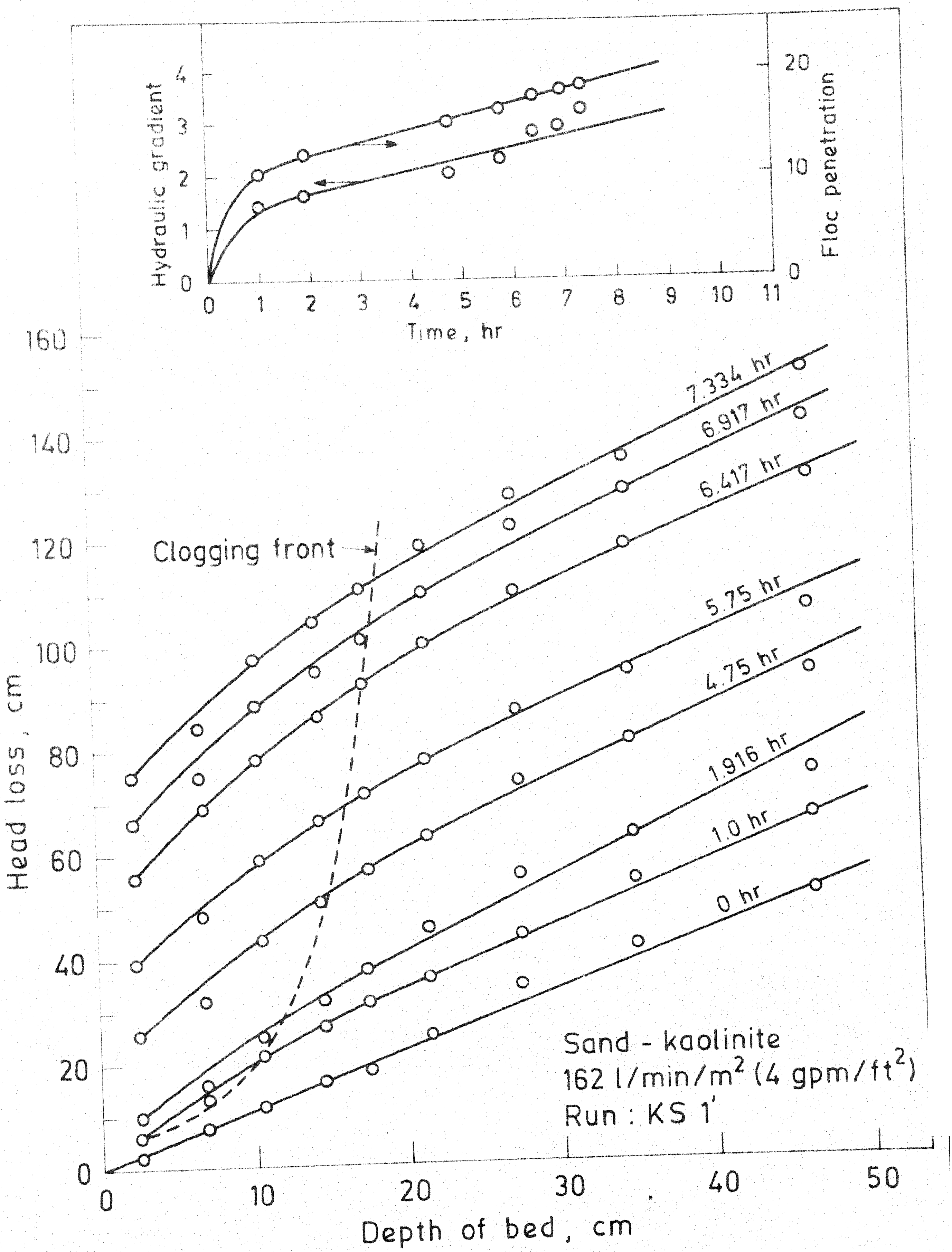


Fig. A1. Headloss hydraulic gradient and floc penetration in sand-kaolinite system at 162 l/min/m^2.

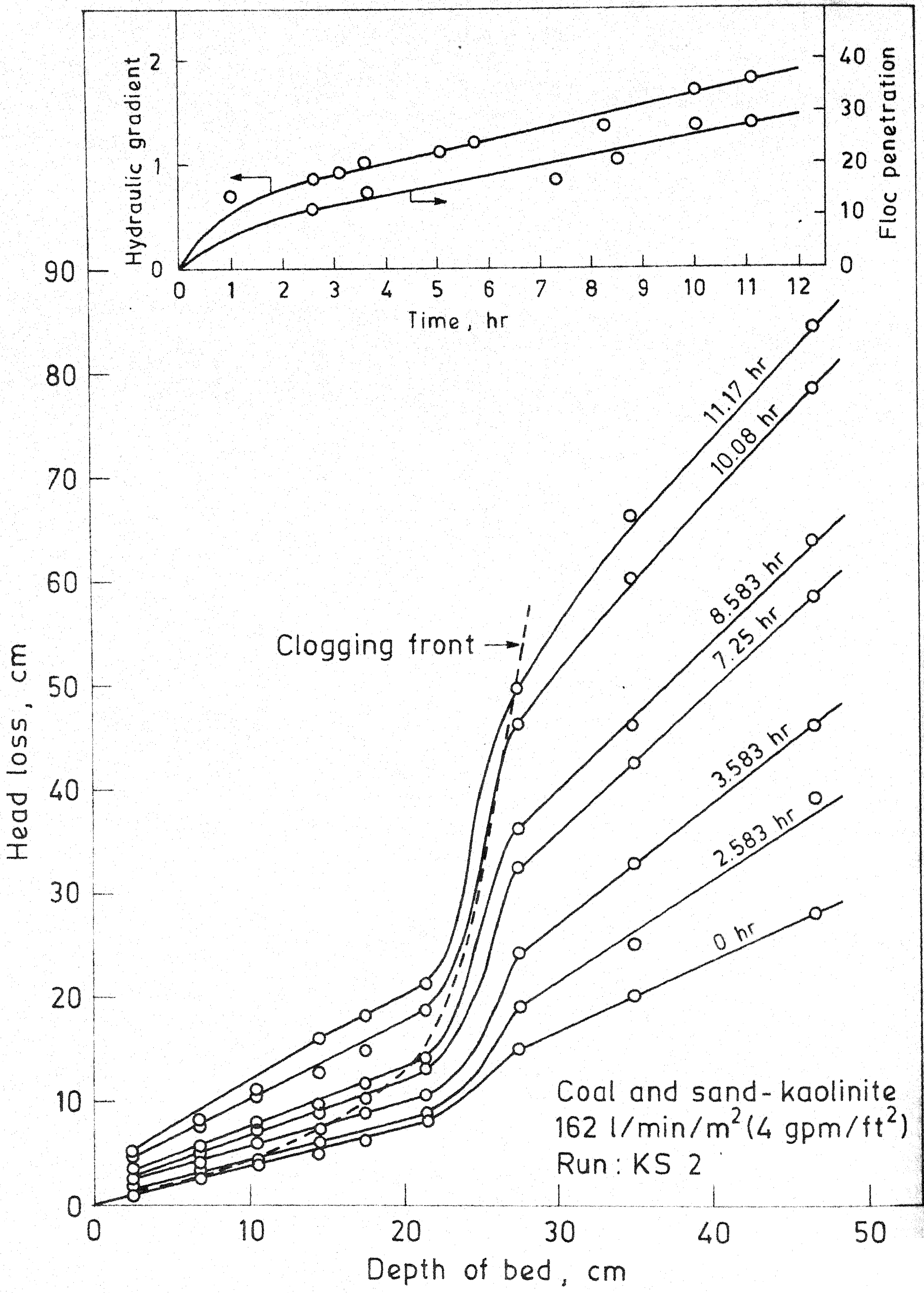


Fig. A2. Head loss, hydraulic gradient and floc penetration in coal and sand-kaolinite system at 162 l/min/m^2.

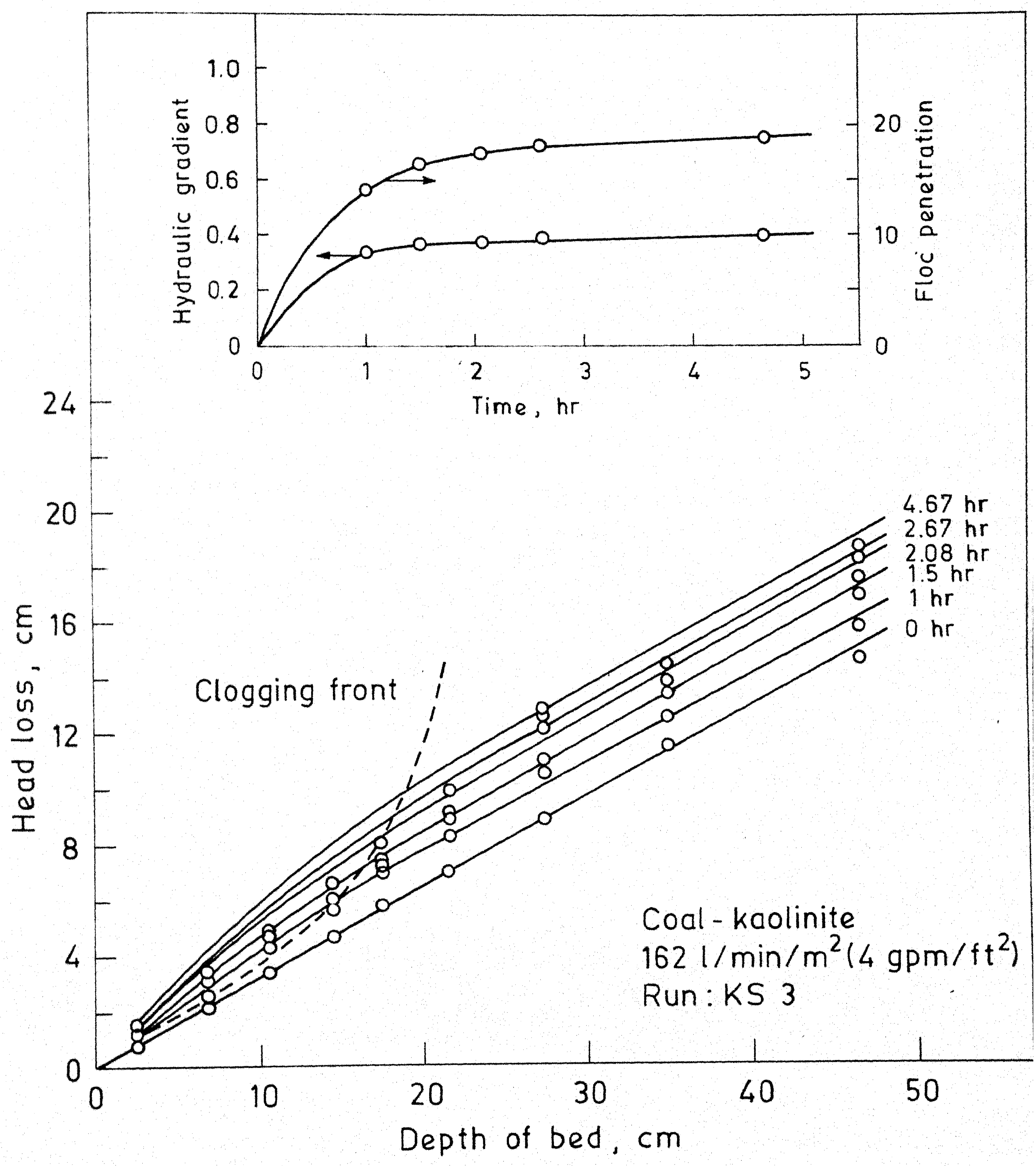


Fig. A3. Head loss, hydraulic gradient and floc penetration in coal - kaolinite system at 162 l/min/m^2.

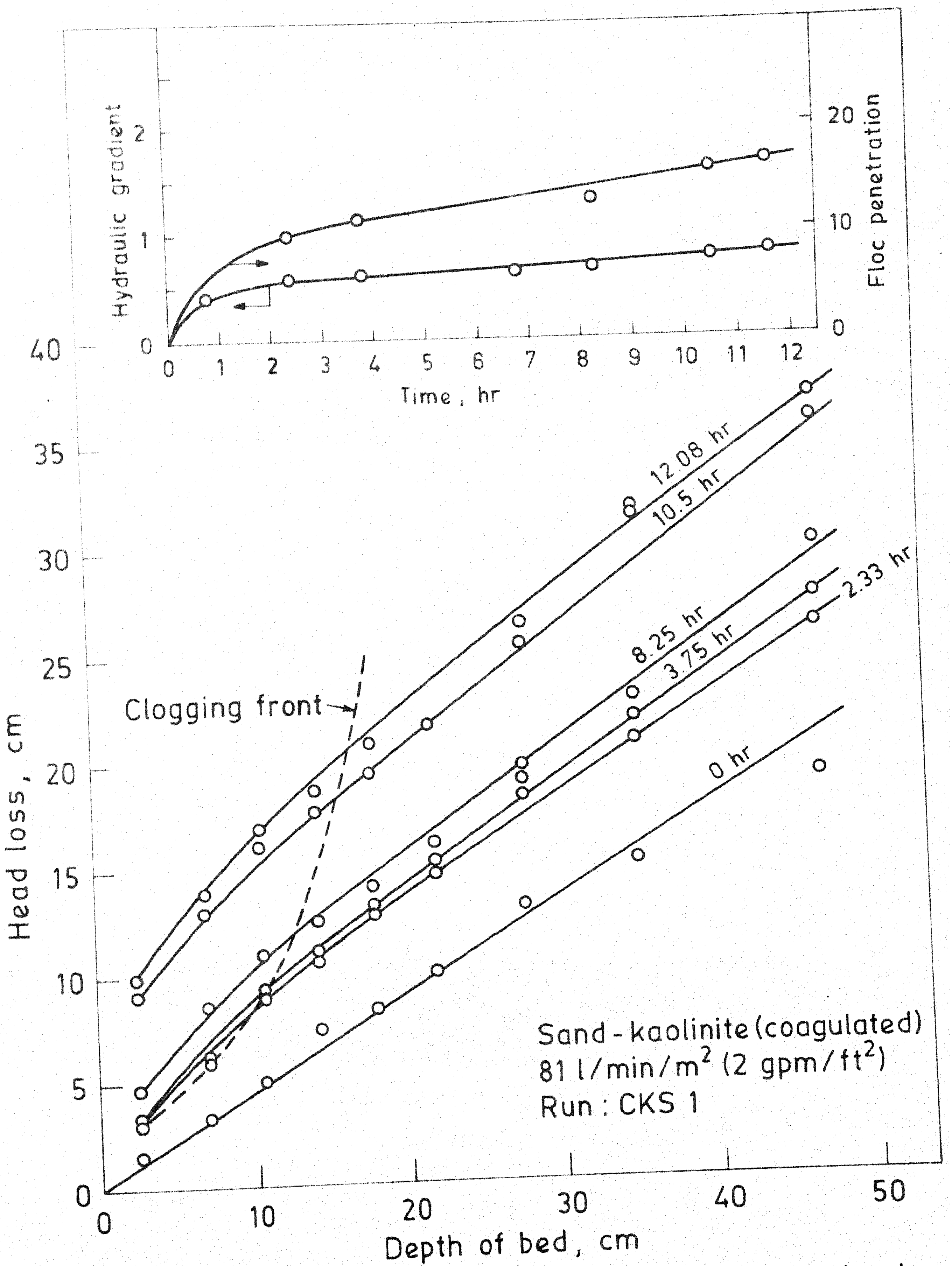


Fig. A4. Head loss, hydraulic gradient and floc penetration in sand - kaolinite (coagulated) system at 81 l/min/m^2.

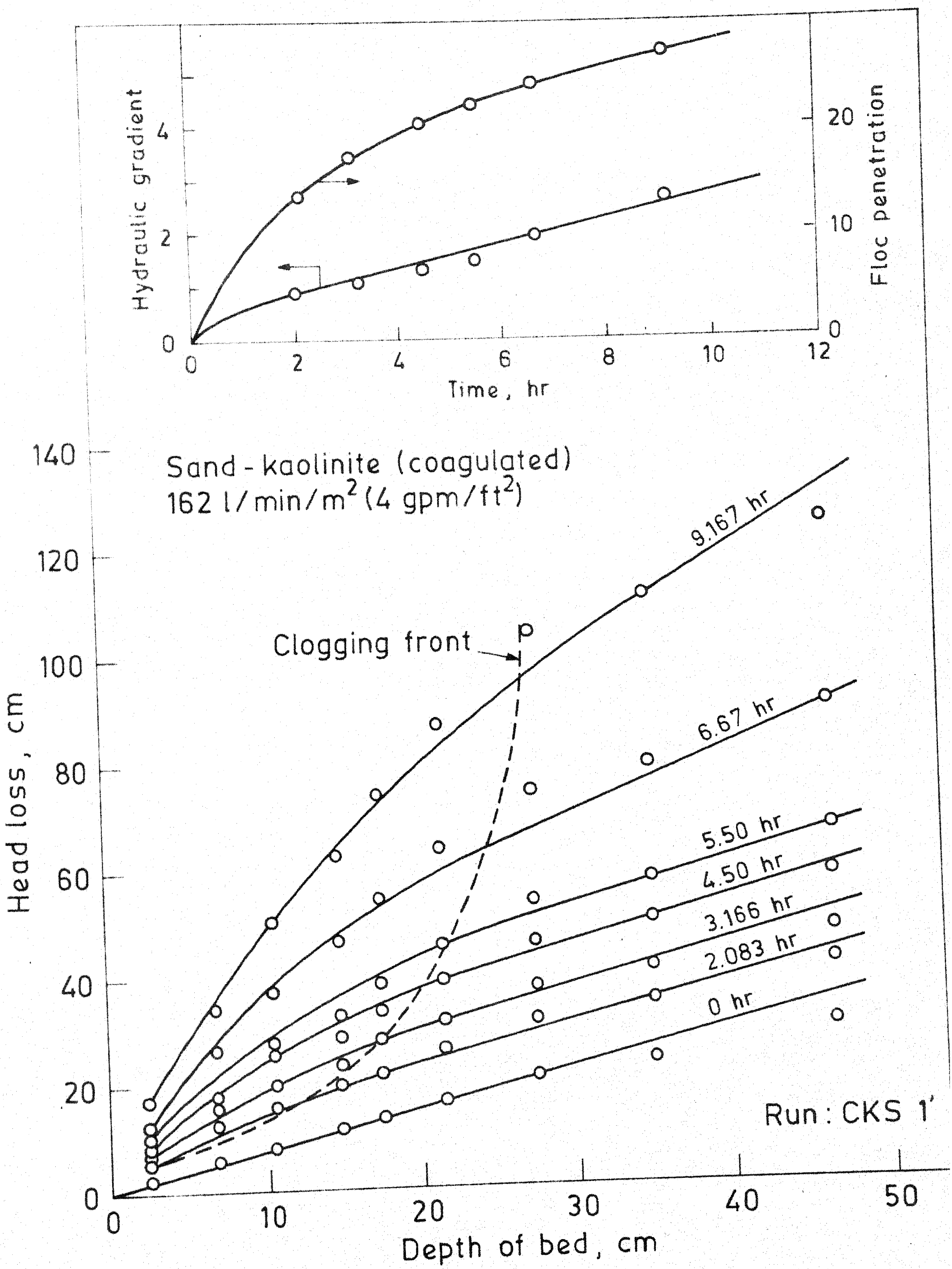


Fig. A5. Head loss, hydraulic gradient and floc penetration in sand - kaolinite (coagulated) system at 162 l/min/m^2.

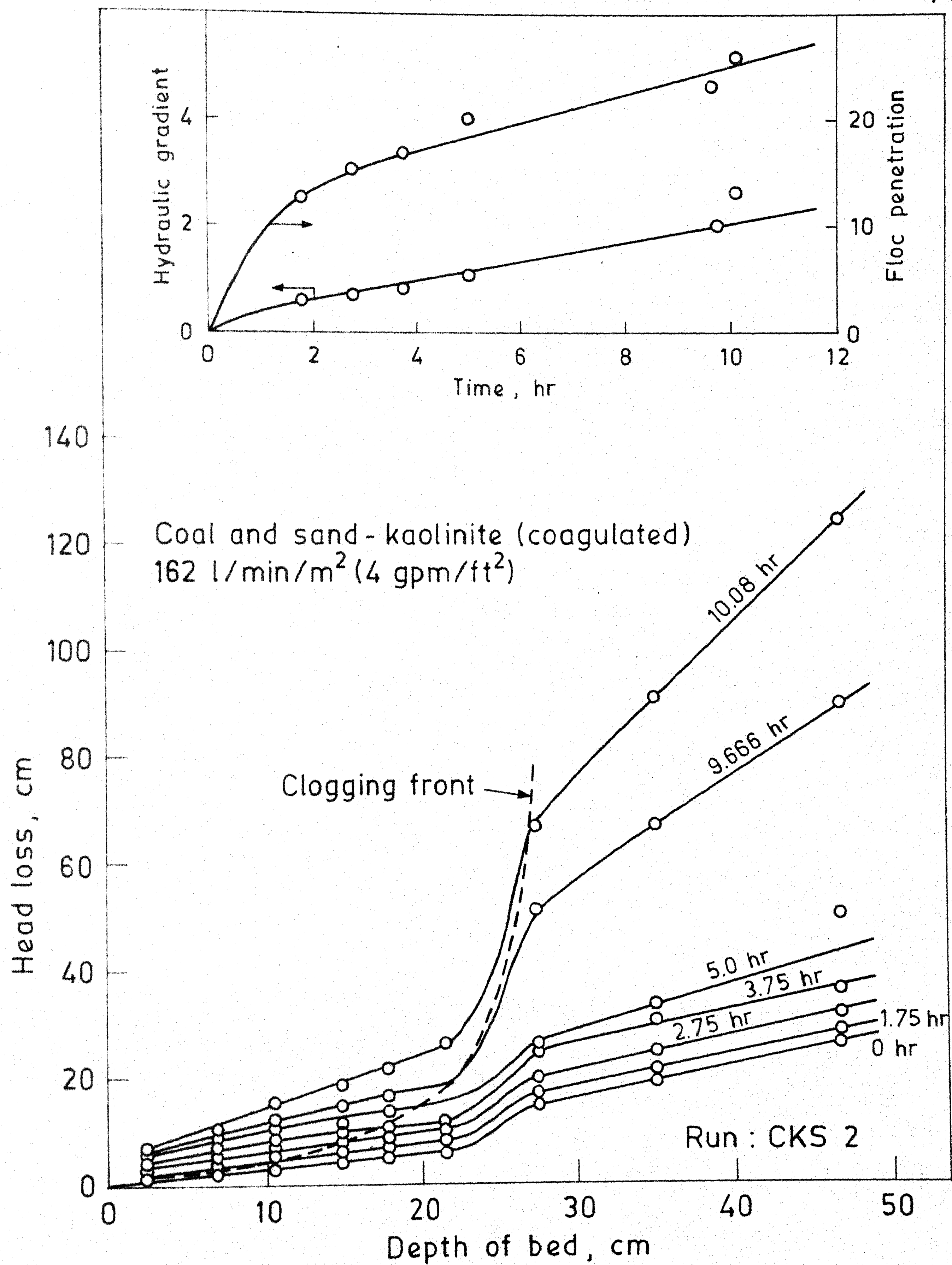


Fig. A6. Head loss, hydraulic gradient and floc penetration in coal and sand-kaolinite (coagulated) system at 162 l/min/

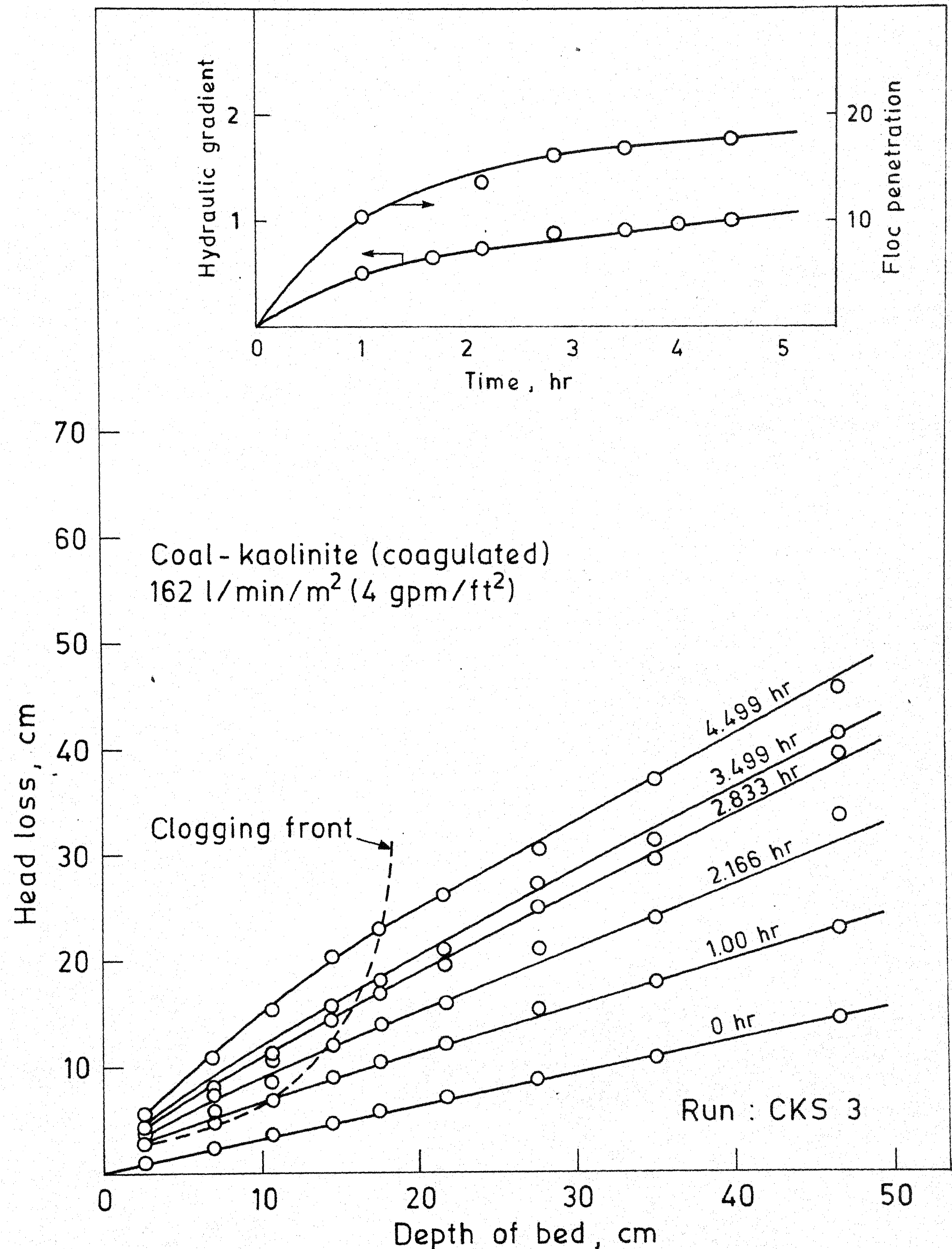


Fig. A7. Head loss, hydraulic gradient and floc penetration in coal-kaolinite (coagulated) system at 162 $l/min/m^2$.

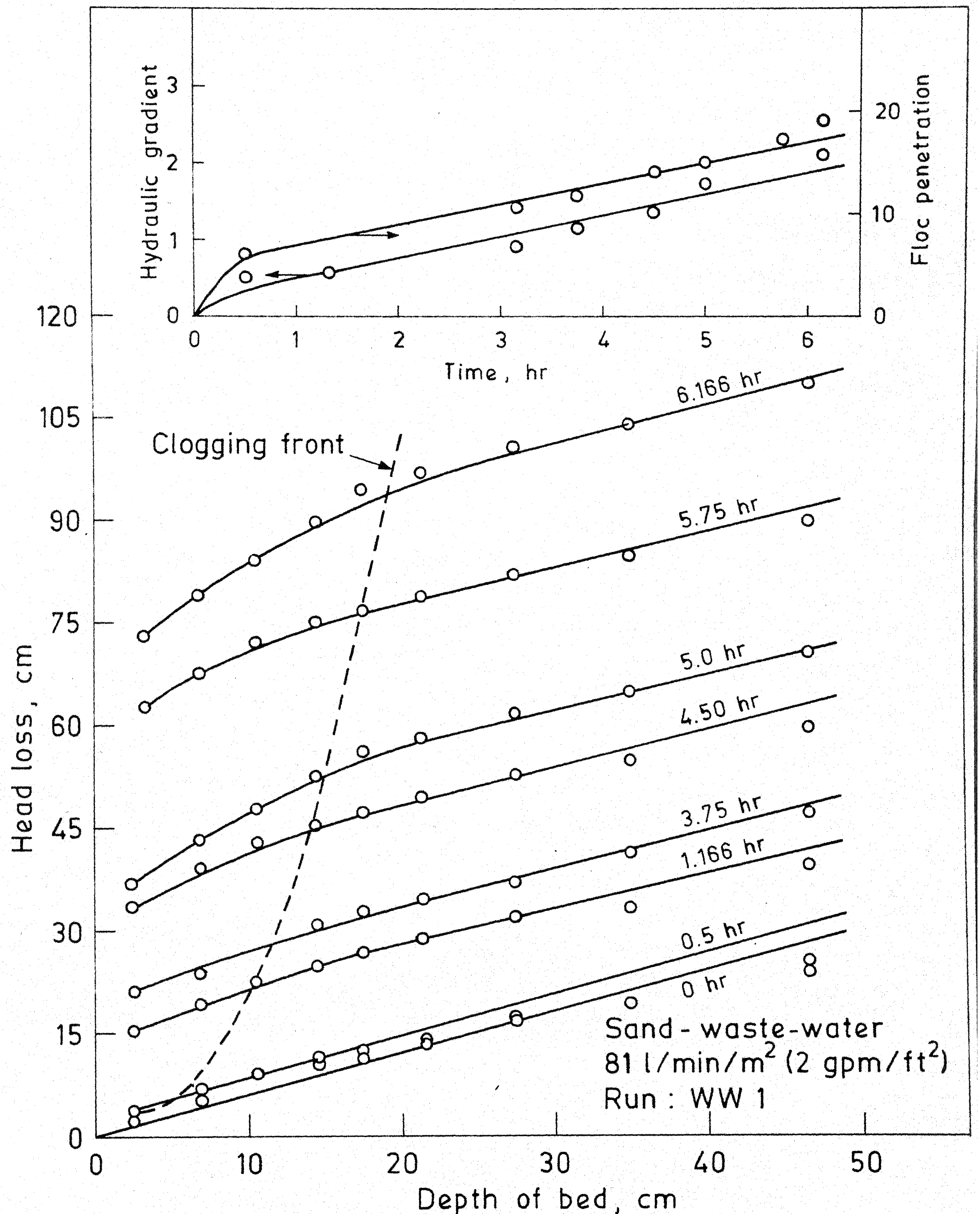


Fig. A8. Head loss, hydraulic gradient and floc penetration in sand - waste-water system at 81 l/min/m^2.

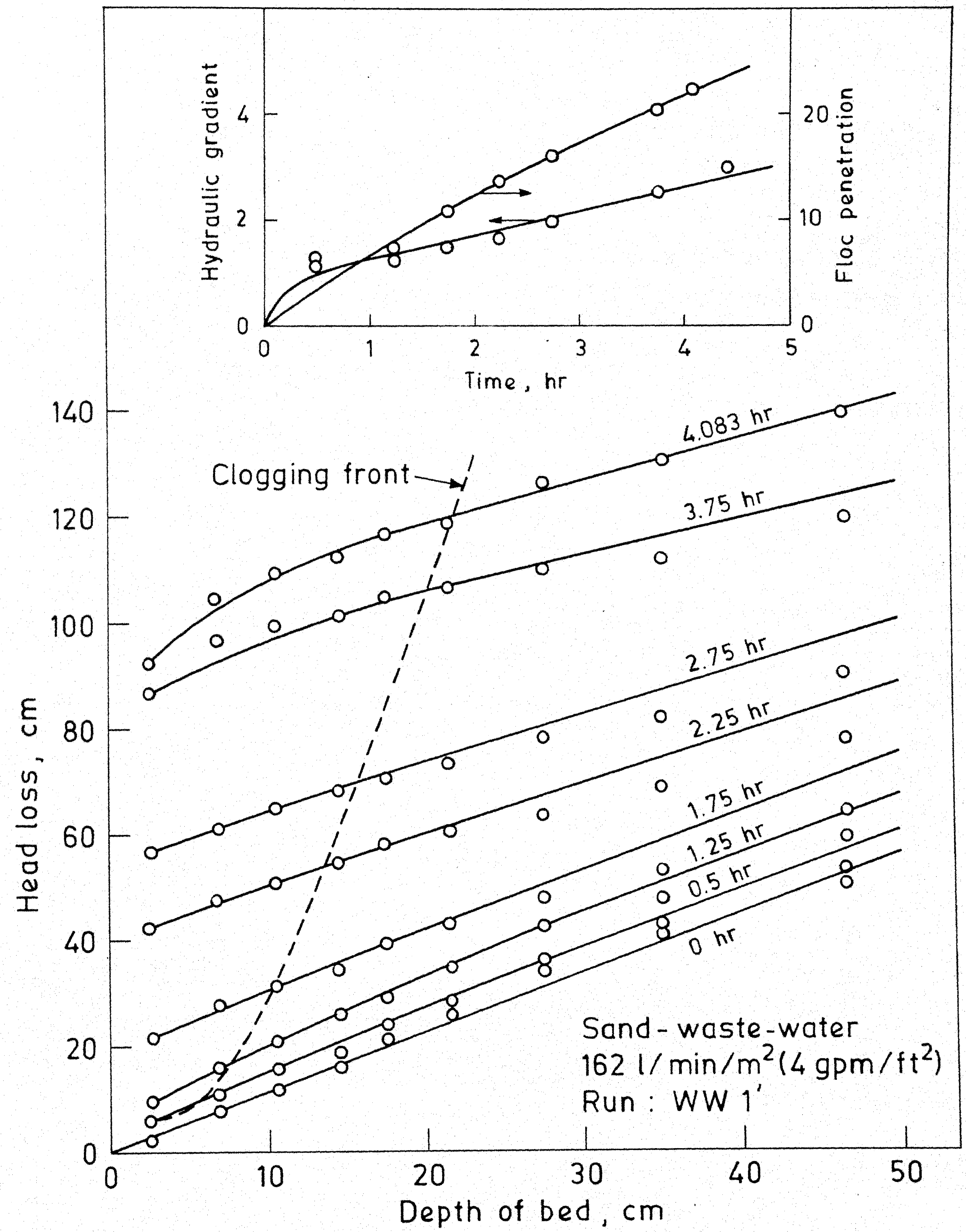


Fig. A9. Head loss, hydraulic gradient and floc penetration in sand - waste-water system at 162 $l/min/m^2$.

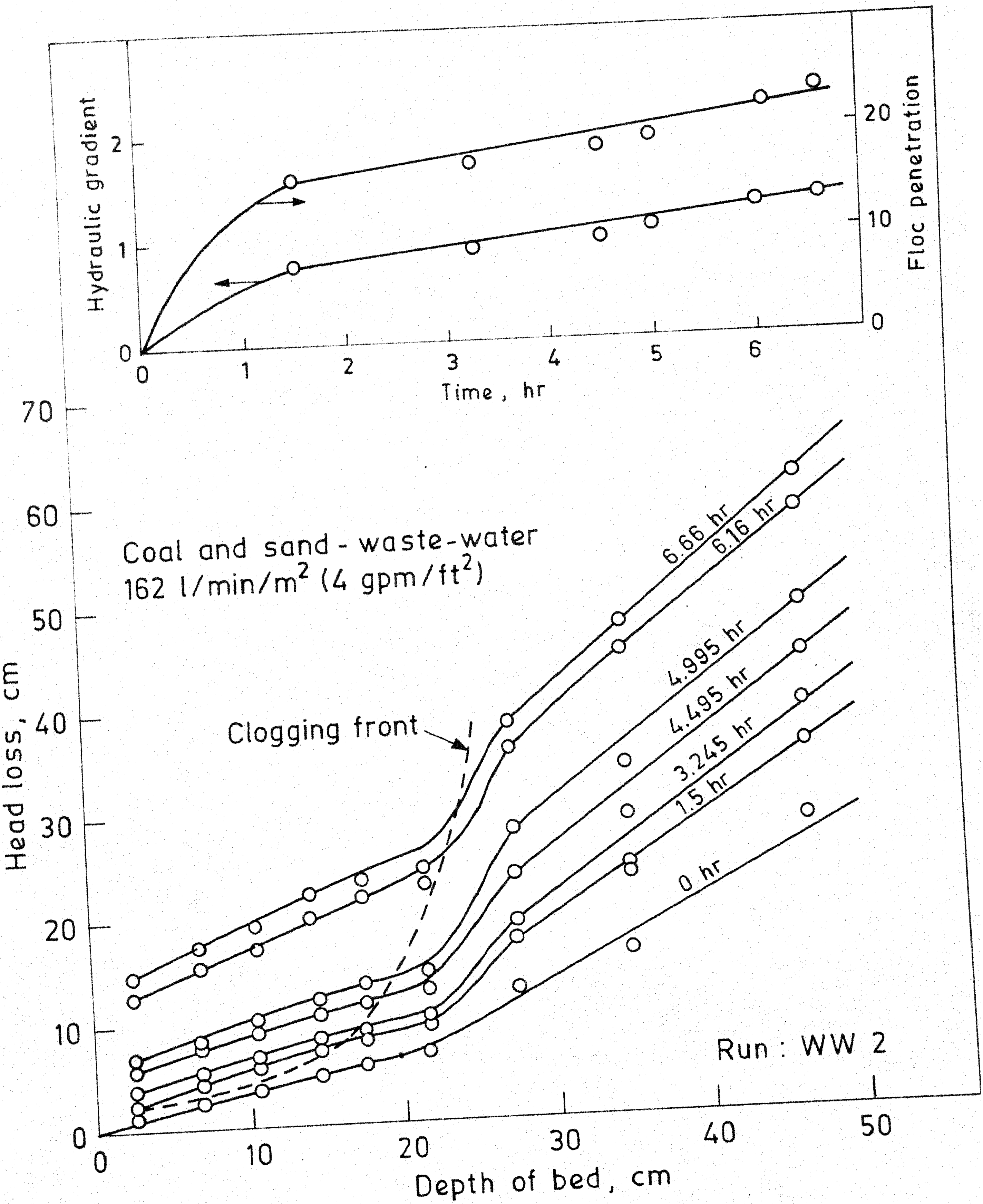


Fig. A10. Head loss, hydraulic gradient and floc penetration in coal and sand-waste-water system at 162 $l/min/m^2$.

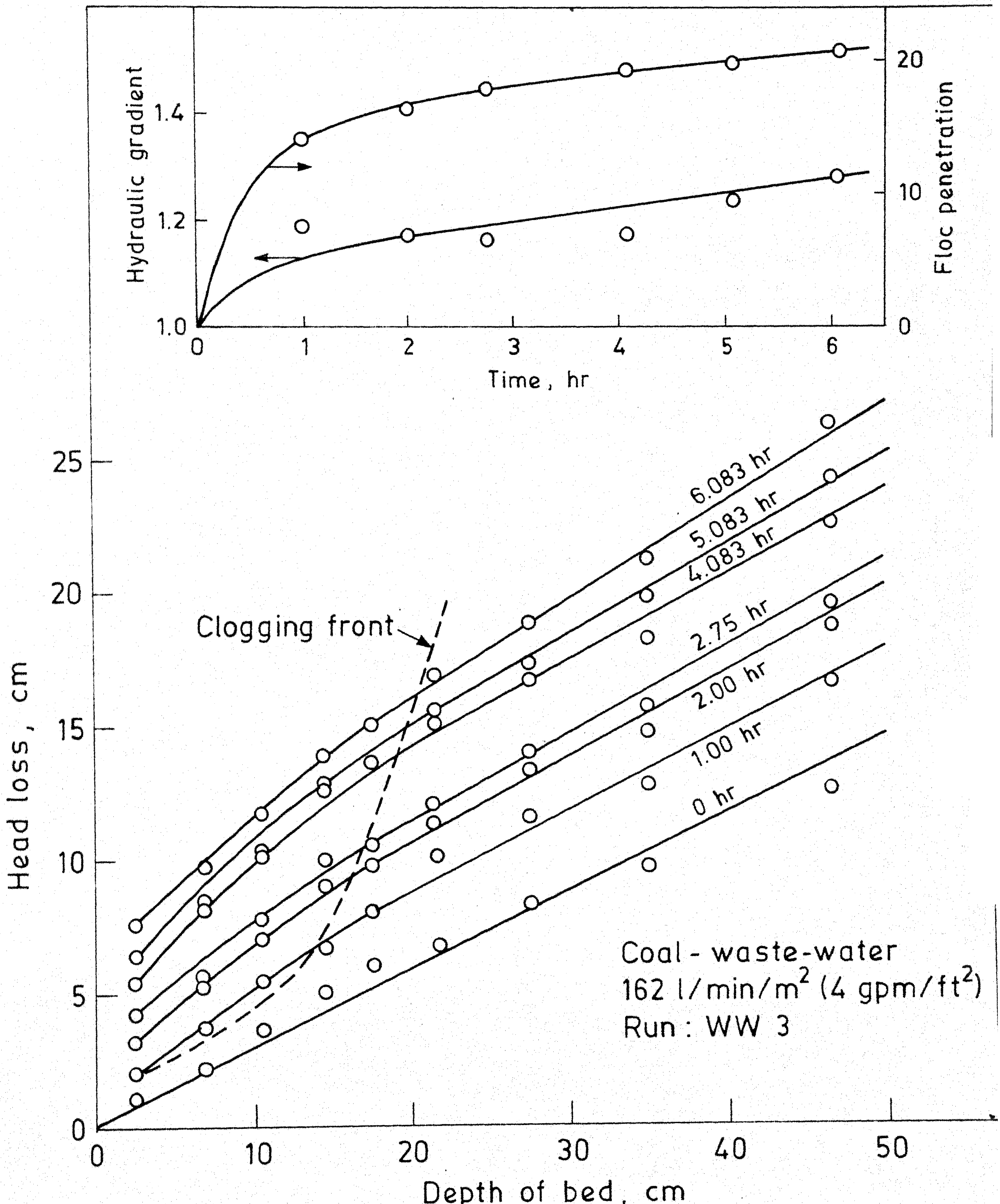
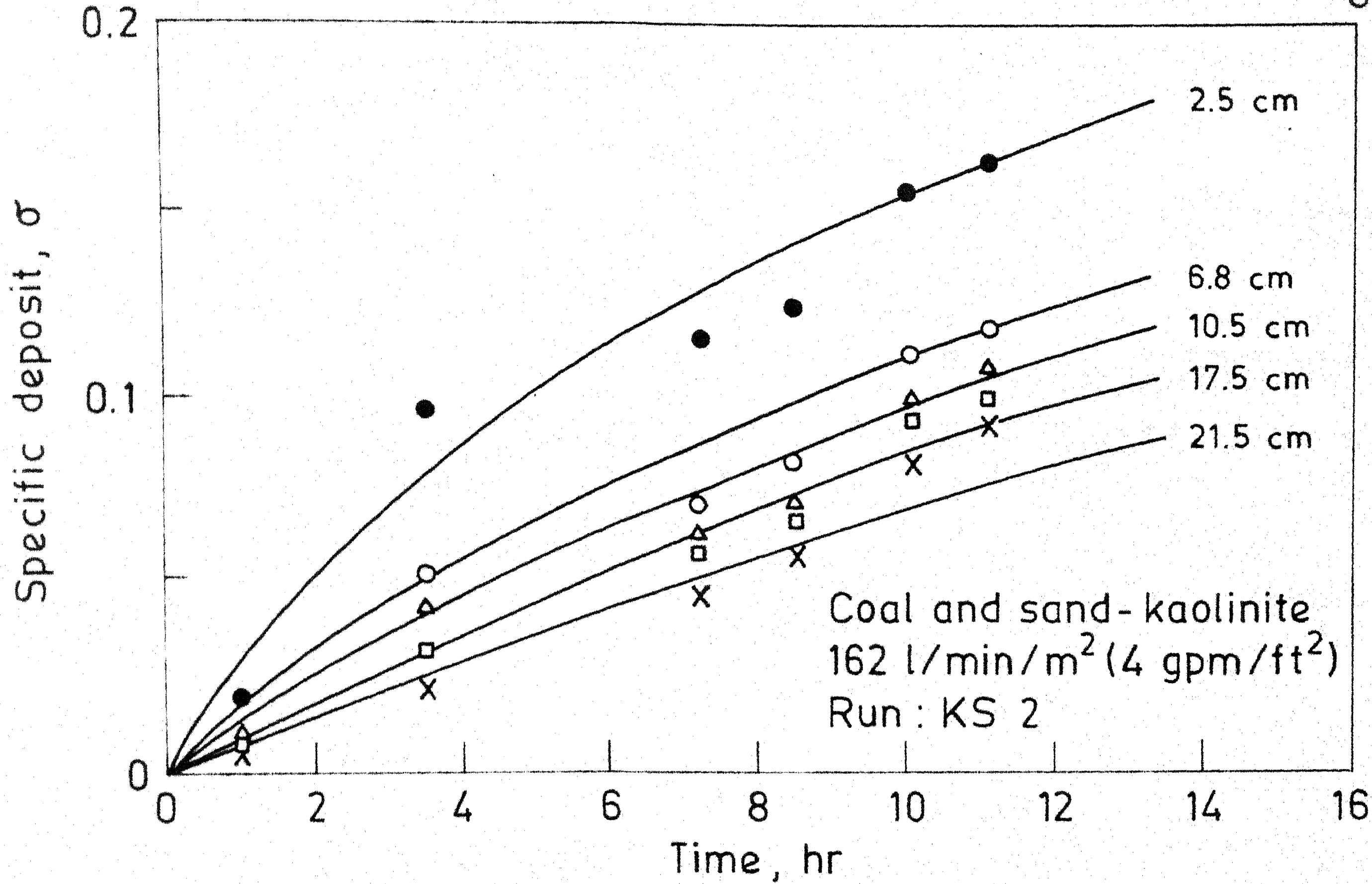


Fig. A11. Head loss, hydraulic gradient and floc penetration in coal-waste-water system at 162 $l/min/m^2$.



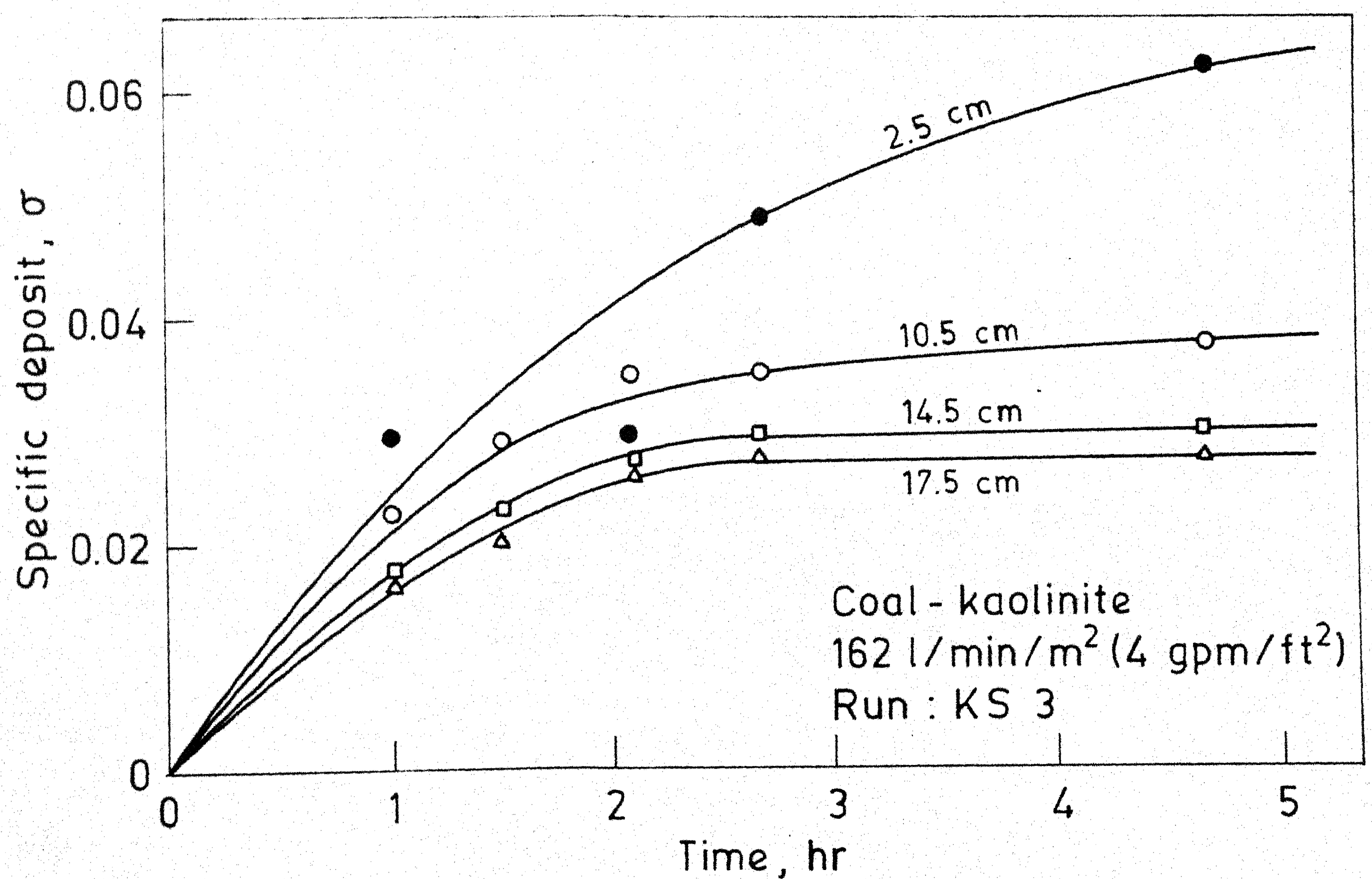


Fig. A12. Specific deposit change during coal and sand, and coal filtration of kaolinite suspension.

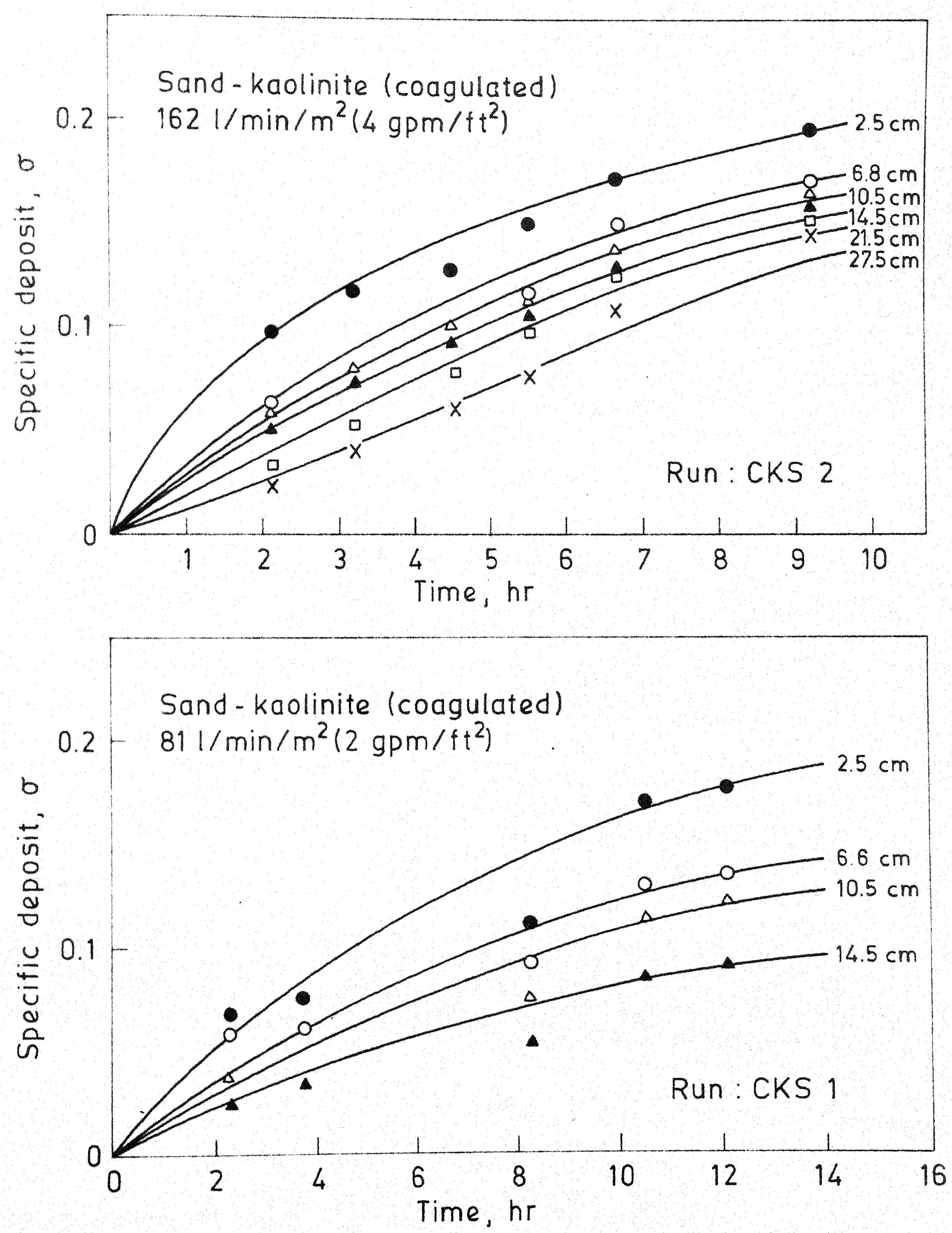
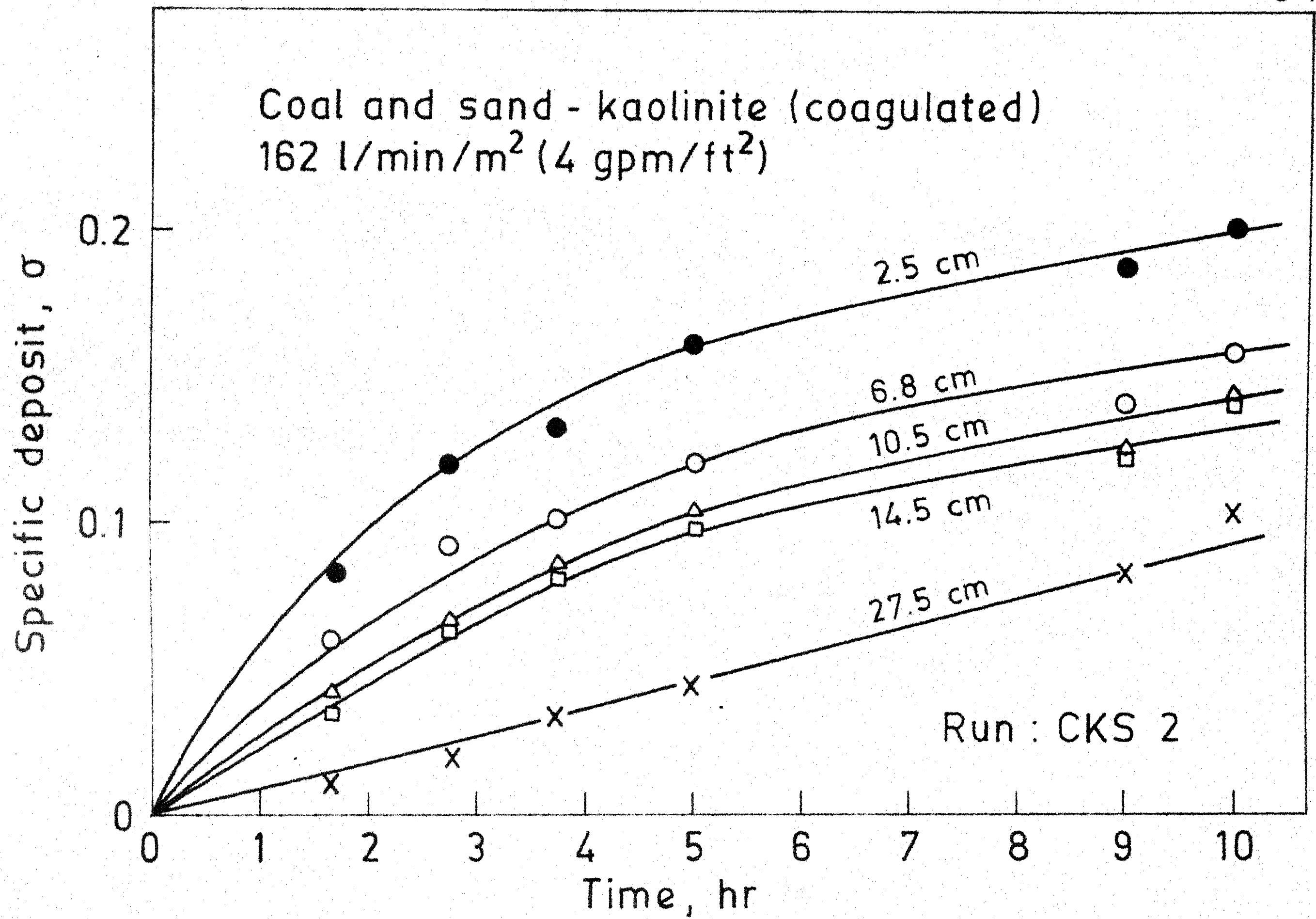


Fig. A13. Specific deposit change during sand filtration of kaolinite (coagulated) suspension



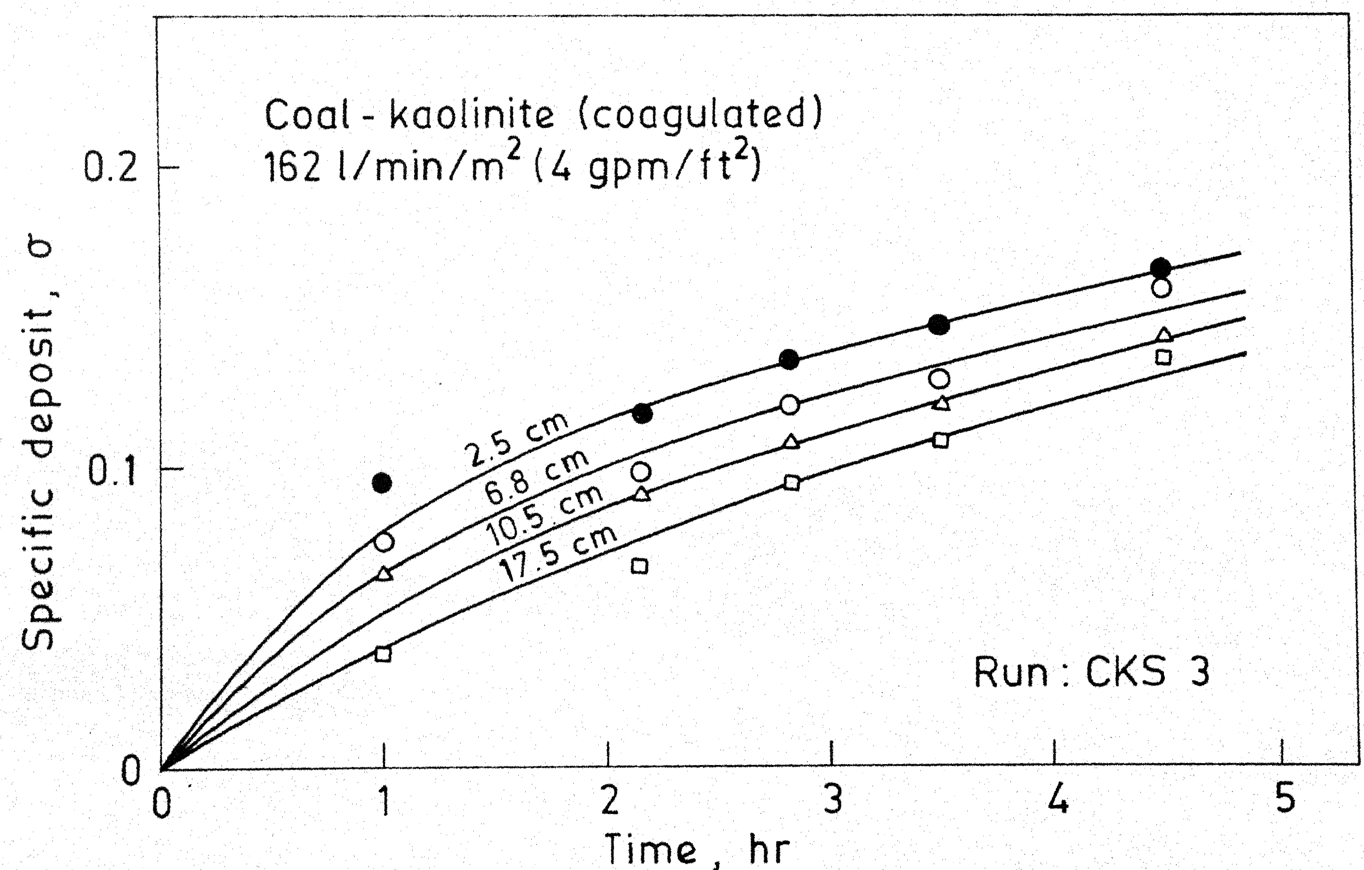


Fig. A14. Specific deposit change during coal and sand, coal filtration of kaolinite (coagulated) suspension.

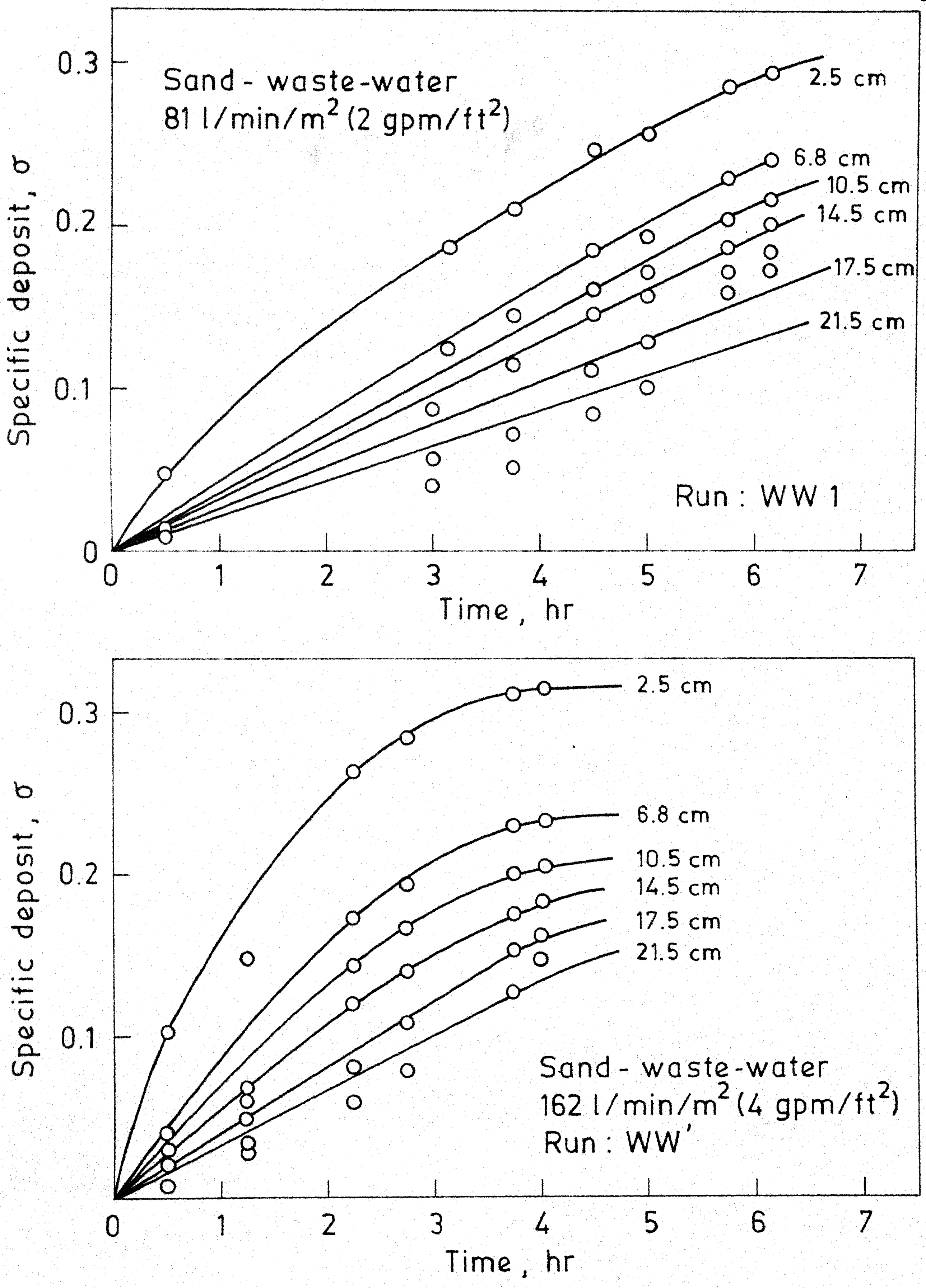


Fig. A15 Specific deposit change during sand filtration of waste-water suspension.

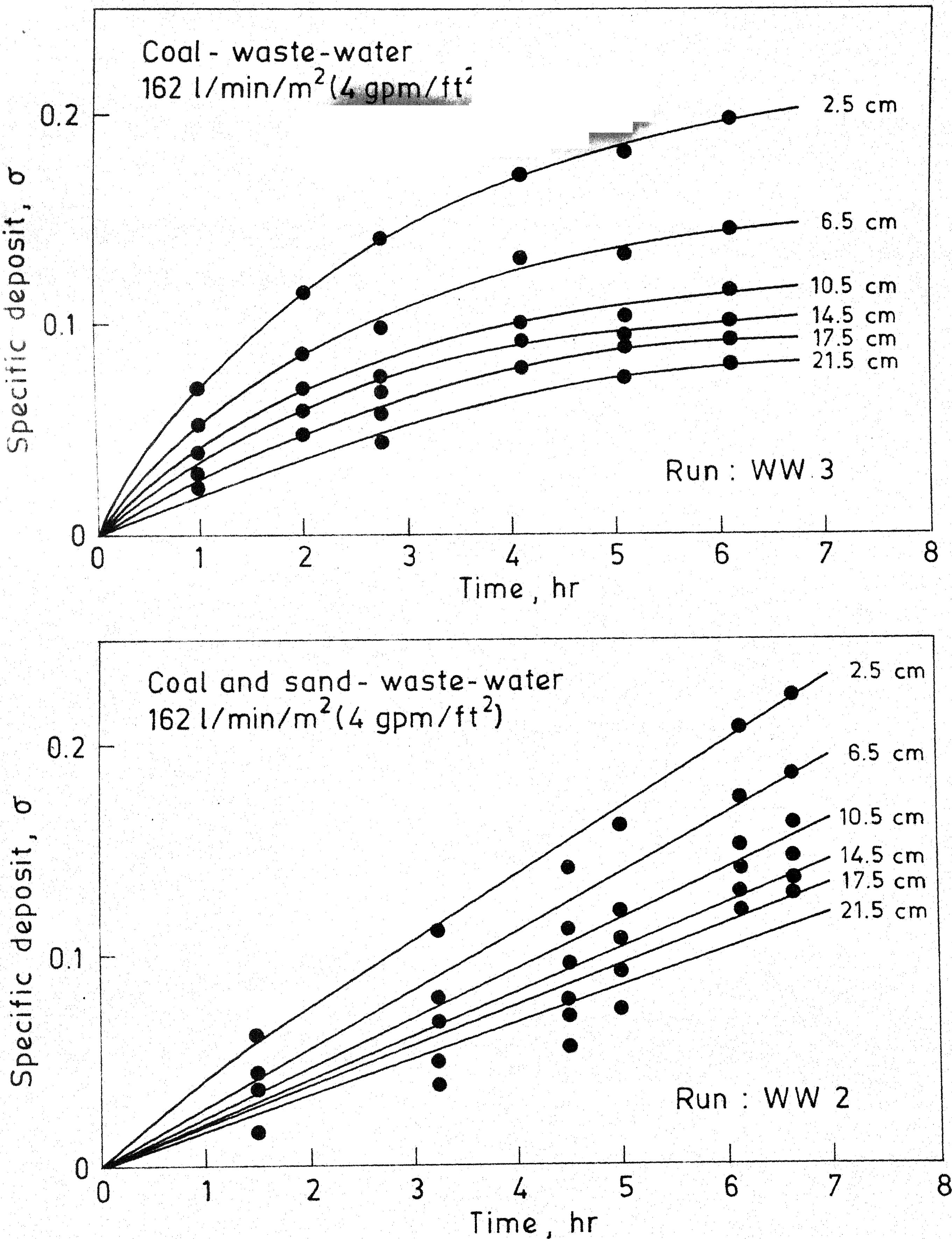


Fig. A16. Specific deposit change during coal and sand, coal waste-water suspension.